Environmental Protection Service
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Study on the Incremental Refinery Construction Costs Associated with the Reduction in the Amount of Lead Additive in Gasoline

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MHG International Ltd.

INCREMENTAL COSTS TO THE REFINING INDUSTRY ASSOCIATED WITH THE REDUCTION IN LEAD ANTIKNOCK USAGE IN GASOLINE

FOR

ENVIRONMENT CANADA

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ABSTRACT

The effect on the Canadian oil refining industry of reducing motor gasoline lead additive content while maintaining adequate antiknock quality was investigated using a computerized process simulator. A composite refinery approach was used in which the refineries in the Atlantic, Quebec, Ontario and Western regions were represented separately in 1979 by a single regional refinery. Capital costs, operating costs and construction manpower availability were estimated for each region for three different levels of lead control.

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EXECUTIVE SUMMARY

The principal objectives of this study were to define the extra processing capacity and to estimate the associated incremental capital and operating costs which the 37 Canadian refineries would incur if lead additive levels in motor gasoline were progressively reduced from 1979 levels to zero, whilst retaining the octane quality.

To reflect the different crude diets and product slates prevalent throughout the Provinces, Canadian refineries were segregated into four regions - Atlantic, Quebec, Ontario and Western. A process simulator (non-LP) was used to compute the process parameters and models were developed representing the four regional composite refineries. Reliance was placed on user interaction to optimize the process operations to attain approximate minimum crude oil throughputs, whilst retaining original product quantities and specifications.

It was deduced hypothetically that no capital investment would be required if 2 gm/IG and 1 gm/IG lead level controls were imposed. However, many individual refineries may detract somewhat from the 'average' refinery configuration used, and may necessitate some process unit modifications or changes in marketing strategy.

To conform to a zero lead level control, the model indicates that significantly more processing capacity would be required in all regions. The approach adopted in the study was one whereby existing process unit capacity would be utilized to the best extent possible and as such, extra process units would be confined to Catalytic Reforming, virgin naphtha (LSR) Isomerization and Alkylation. The associated incremental capital costs have been estimated to be approximately $886 million Canadian on a 1979 basis.

The regional cost split is: -

Atlantic	20 %
Quebec	23
Ontario	29
Western	28

Product property specifications could be met with relative ease but the potential difficulties in the Quebec and Atlantic Regions are currently obviated by use of MMT to produce unleaded gasolines. Crude oil increments corresponding to the zero lead case are listed below, together with corresponding energy penalties: -

	Crude Oil Increment	Energy Penalty
	%	%
Atlantic	1.05	0.09
Quebec	1.51	0.34
Ontario	1.16	0.40
Western	2.26	0.82

Incremental operating costs were calculated based on the converted regular motor gasoline pool volume only in 1979 in each region and the results are shown below: -

	Canadian cents/litre		
		Control Level	
	2gm/IG	1gm/IG	No Lead
Atlantic	0.20	0.43	2.52
Quebec	0.17	0.69	1.80
Ontario	-	0.03	1.37
Western	-	0	1.05

Projected construction schedules have been developed and a staggered four to five year program is recommended. Published future projects in the refinery and petrochemical industries will create constraints through skilled-trade construction labor shortages, and pipefitters are expected to be especially critical in the Western region.

It is considered that there will be many opportunities for refineries to specialize their markets to alleviate some of the constraining parameters induced when lead-free gasoline is produced. Essentially, they are categorized as 'premium gasoline' and ' diesel fuel' refineries. Refineries in remote locations are envisaged to incur the highest cost penalty to produce lead-free gasoline.

Regional variations in the gasoline/distillate ratio are marked and range from 0.64 in Atlantic to 1.21 in Western region. This parameter in addition to the crude diet should be appreciated when evaluating the regional differences.

Brief discussions are included in the report on the following topics:

- future gasoline production to 1995
- heavy oil refining
- technological progress to produce high octane gasoline without the use of lead anti-knock additives.

The process capacity and cost estimates produced in this report should not be regarded as definitive,but should be interpreted in the light of serving to illustrate the directional trends which refiners might assess when demands for increasing proportions of lead-free gasoline are encountered. Also the study indicates potential strategies which refiners might adopt to alleviate known constraints.

SECTION 1

1.0 INTRODUCTION

This report has been prepared for Environment Canada to assist in their assessment of the socio-economic impact of reducing the amount of lead antiknock additive in motor gasoline in Canada. Incremental capital and operating costs which would be incurred by the petroleum refining industry, have been investigated in addition to the associated construction labour demand/availability projections.

During 1979, thirteen companies operated 37 petroleum refineries in Canada and produced approximately 235 million barrels of gasoline of which approximately 68% contained lead antiknock additive. The quantity of lead used amounted to some 26 Million pounds.

Originally, it had been intended that the study examine the effect upon individual refineries in order to obtain as accurate an assessment as possible; however, such an analysis required a large amount of detailed information, much of which the refineries considered confidential and time consuming and which they were reluctant to release for this study. Accordingly, it was decided with Environment Canada's agreement that the study would be restricted to a general examination of the petroleum refining industry on a regional basis, with the results so obtained proportioned out to each individual refinery in the region.

The specific objectives of the study were to estimate the capital and operating cost impact on Canadian refineries based on the 1979 scenario, resulting from three different levels of lead control. Accordingly, the following four cases were investigated using a computerized general refinery model:

Case 1: Base case simulations reflecting 1979 Refinery operation
Case 2: Reduction in lead additives to 2 grams per imperial gallon

Case 3: Reduction in lead additives to 1 gram per imperial gallon

Case 4: Complete ban on lead additives to gasoline.

SECTION 2

2.0 METHODOLOGY

2.1 DESCRIPTION AND APPLICATION OF COMPUTER MODEL

Owing to significant differences in crude oil diet used by refineries, it was decided to divide Canada into four distinct regions, namely Western, Ontario, Quebec and Atlantic. Figure 2-3 defines these regions geographically.

A computer model was developed to reflect the general refinery process unit configuration shown in Figure 2-1. Refinery operations in the four major regions of Canada were simulated by composite refineries representing each region. The capacity of each process unit in each of the four composite refineries was taken to be equal to the sum of the installed capacities in 1979 of all the units within that region.

The computer model used is a process simulator and provides for material balance and product property calculation. No economic data was incorporated into the program but reliance was placed on user interaction to develop an optimum solution.

In preparing study cases, the non-LP model was solved interatively by means of user interaction. In effect the optimizing procedure such as would be performed by a linear programming algorithm was done manually. The optimizing criteria applied for each case study were:-

a) Maximum utilization of existing process units.
b) Minimum crude oil consumption

Incremental crude oil consumption is disadvantageous to Canada both economically and politically and thus was considered to be the prime factor in the determination of the optimal solution.

A fairly systematic approach was used to develop each case. The steps followed are illustrated in Figure 2-2 and are summarized below:-

1. Data gathering - each case study requires a definition of feedstock diets, product slates and unit capacities. These are provided for each region in Tables 2-1 to 2-12.

2. Execution of crude blending routine (blocks 1-3) in Figure 2 - 2) - this computer routine breaks down the individual crudes into pseudocomponents and blends them in the specified proportions.

3. Execution of material balance routine (blocks 4 - 10) - sufficient data are entered to close the material balance, such as product slates, feed rates to units, some TBP cut points, cat cracker conversion and characterization factor of the naphtha feed. The program then executes to estimate yields from the process units and to complete the material balance, providing flow rates for all the streams shown in Figure 2 - 1.

4. User inspection of material balance results (block 11 in Figure 2 - 2) - The results are reviewed against the following constraints:-

    a) Unit capacities (Tables 2-9 to 2-12)
    b) Limitations on certain operating variables such as cut points, unit turndowns (Table 2-13)

    If these constraints are not satisfied, step 3 is repeated with modified input data.

5. Execution of product quality routine (blocks 12 and 13) - additional data (including gasoline lead concentration) is entered by the user and product qualities are then computed.

6. User inspection of product qualities (block 14) - the estimated product qualities are reviewed against the product specifications listed in Table 2-14. If the specifications are not satisfied, the analysis is returned to either step 3 or 5 as desired.

7. User review of optimizing criteria - if a solution is found to satisfy all the constraints, it is then reviewed to ensure that it is more or less optimal in terms of the optimizing criteria mentioned above. In general, cases were repeated in the optimal direction until one or more slacks between a variable and its constraints were eliminated.

Thus, 'Base Cases' were developed for each region which reflected the 1979 refinery operations in terms of crude diet, product quantities, lead additive concentration, etc.. Subsequently, basic process unit parameters such as severity level and throughput of the catalytic reformer and FCC units were varied to satisfy all the product property specifications, consistent with effecting approximate minimum crude oil throughput. The crude oil throughput so calculated had to agree, within a close tolerance, with the published crude oil consumption figure for that region in 1979. Table 2-20 illustrates the approximate crude oil yield slate applicable to each region. Table 2-21 illustrates the regional differences in gasoline consumption during 1979. A leaded regular gasoline octane quality of 93.1 RON was used to develop all the regional Base Case and a minimum reformer severity of 98 RON was necessary to satisfy all constraints.

For each region, cases were developed to conform to the reduced lead levels in regular gasoline of 2 gms Pb/Imperial Gallon and 1 gm Pb/Imperial Gallon, whilst maintaining the octane quality constant at 93.6 RON. The extra octane barrels were derived principally by increasing the throughput and severity of the catalytic reformer and utilizing other existing process unit capacity optimally.

To produce a totally lead-free gasoline pool, the 'leaded regular' portion was adjusted to have an octane quality somewhat less than the un-leaded regular grade as was already produced, (93.6 RON), to allow for the use of MMT. Similarly, the unleaded premium gasoline octane number was downrated. The octane numbers used depend on region and are referenced in Table 2-22. An isomerization unit and increased alkylation unit capacity was included in the model for the no-lead cases only.

For all cases, the original product slate was maintained with the exception of fuel gas which floated. When product properties could not be met, the reformer severity was increased until these constraints were satisfied.

The constraints which had the greatest effect on the flexibility are listed below:-

- Research octane number (RON) of the gasoline pool in conjunction with reformer capacity.
- Light straight run (LSR) and naphtha capacity minimum TBP end point limitations.
- Heavy Fuel oil maximum allowable viscosity.
- Distillate product specifications such as diesel fuel cetane index and pour point and light fuel oil 50% ASTM distillation temperature.

After completing the above procedure, capital cost and energy penalties were estimated manually for each case.

2.0 METHODOLOGY

2.2 BASIC PARAMETERS

Crude Diets (Tables 2-1 to 2-4)

The feedstock diets for 1979 were obtained from Statistics Canada (Reference 8) and National Energy Board data (Reference 7). For simplification, feedstocks from only the major sources (Saudia Arabia, Venezuela and Iran) were considered. Breakdowns for light and heavy Arabian and Venezuelan crudes are approximate and were obtained by consultation with EMR. All four of the major domestic feedstock components - light, heavy synthetic and condensate/pentane plus were included.

The individual crudes listed in Tables 2-1 to 2-4 are actually blends of several crudes from different fields. The specific assays used to represent each blend are summarized in Table 2-15. Assays for the domestic crudes were taken from in-house reference files while assays for the imported crudes were taken from Oil and Gas Journal (Reference 9 and 10).

Two values of sulphur content are given for domestic light in Table 2-15. The higher value was used for the Western Region and is based on sulphur production figures in that region (Reference 11).

Product Slates (Tables 2-5 to 2-8)

Data reported as "Refinery Output" by Statistics Canada (Reference 8) were used as 1979 production figures for each region. Refinery output is reported for total gasoline rather than for each grade. The production of each grade was therefore taken to be proportional to net sales data which are also available in the same reference.

Additional information was required to estimate the relative amounts of various types of petrochemical feedstocks and this was done based on the capacities of petrochemical plants as published in Oilweek (Reference 12).

Refinery Configurations(Tables 2-9 to 2-12)

Capacities for the major processing units installed in each refinery as of the end of 1979 were obtained from an EMR publication (Reference 1).

This data also identified the hydroskimming refineries within Canada, i.e. those without vacuum units. The operation of these refineries were represented by fixing the flow of reduced crude to heavy fuel oil from the atmospheric crude units (stream 31 in Figure 2 - 1) to some estimated value. In the Atlantic region, this stream was estimated to be 40000 BPCD based on the following contributions:-

Gulf - Point Tupper:	20,000 BPCD
Ultramar - Holyrood:	5,000 BPCD
Irving - Saint John:	15,000 BPCD

The allowance for the Irving refinery is included because the reported vacuum unit capacity (65,000 BPCD) is small relative to the total crude capacity.

In Ontario, a reduced crude rate to heavy fuel oil of 54,000 BPCD was taken as the total contribution from the Petrosar operation and the Suncor topping refinery.

Refinery Operating Parameters (Table 2-13)

These constraints were based on in-house experience and reference files pertaining to refinery design and operation.

Product Specifications (Table 2-14)

The number of product types was minimized and their specifications were derived using an average of the specifications for the range of grades of products within a particular broad category.

The specifications published by the Canadian Government Standards Board were supplemented by additional data on product quality attained in practice and appropriate target levels which were available from both in-house references and the major refiners themselves.

The same specifications were adopted in a recent similar study performed by MHG for Transport Canada to assess future availability of transport fuels.

Gasoline octane numbers (RON) for the three basic grades were derived from a recent independent study on Canadian gasoline sales (Reference 2). Extent of MMT use was applied by judgement.

Process Yields

Catalytic reformer, catalytic cracker and isomerization yields were taken from "Petroleum Refining". Technology and Economics (Reference 5). In general these correlations are based on feedstock gravity and appropriate characterization factors derived from the crude assay data.

The following assumptions were made regarding process unit type and appropriate data was incorporated into the computer model:-

Catalytic Reformer	- Mono-metallic platinum Catalyst, semi-regenerative unit
Catalytic Cracker	- FCC unit using zeolite catalyst and heavy vacuum gas-oil feedstock
Alkylation	- HF catalyst using C_3/C_4 olefinic feedstock plus i-C_4.
Isomerization	- LSR (C_5-C_6) feedstock with no recycle.

For the HF Alkylation Unit, iso-butane requirements were taken to be 1.24 volumes per volume of olefin, and iso-butane could be imported as required to react all olefins allocated to alkylation.

It was assumed that the Gas Plant in each composite refinery has the capability to recover 95% iso-butane for alkylation and total C_4's for blending as required from all sources (crude tower overhead, catalytic reforming, catalytic cracking).

For the Distillate Desulphurization Units, sulphur removal was taken as 100% from naphtha/LSR, 95% from kerosine and 90% from light gas oil and light cycle oil. All feed was assumed to be recovered as product with no generation of light ends or naphtha. Product properties were assumed to be the same as the feed properties. Hydrogen consumption estimates were taken from Hydrocarbon Processing (August, 1980). The consumptions used for the various streams are as follows:-

Naphtha/LSR	:	30 SCF/bbl
Kerosine	:	200 SCF/bbl
Light Gas Oil and Light Cycle Oil	:	700 SCF/bbl

Operating Costs (Table 2-17)

The 1979 costs for feedstocks and products were obtained from current publications such as Canadian Petroleum. Utility costs were taken from a recent study performed by Montreal Engineering Ltd..

Energy Penalty

The energy penalty for producing a fixed quantity of lead-free gasoline is defined by Environment Canada as the additional crude processed plus additional purchased power minus the energy equivalent of any additional liquified petroleum gas or other product, all divided by the quantity of crude oil required to produce an equal fixed quantity of leaded gasoline.

Table 2-18 which lists the unit energy requirement for process unit operations, was applied in the calculations. This data was derived from a literature search and a representative number assigned for each process unit considered.

Energy equivalents of utilities which were applied are listed below:-

Electricity	10000	BTU/KW
Steam	1300	BTU/LB
Fuel		HHV (N.B. The basis for FOE of Fuel Gas Product in the computer model is 6.35×10^6 BTU/BBL).
Heavy Fuel Oil (#6)		HHV calculated for each case by formula.
Crude Oil		HHV calculated by formula for each region. (e.g. 165158 BTU/I.G. for Quebec).

The formula used to determine the HHV of crude oil and heavy fuel oils is given below:-

$$BTU/I.G. = (18320 + 40\ (API - 10)) \times (10 \times S.G.)$$

The Heavy Fuel Oil Product in the computer model product slate includes own refinery consumption for fuel.

Product Quality Estimation

The pseudocomponent approach was used to estimate the properties of distillates, turbo fuels and heavy fuel oil. The true boiling point curve for each crude was subdivided into pseudocomponents of 20^{0}F boiling range. Each pseudocomponent was assigned a specific gravity, mid boiling point and sulphur content as specified by the crude assay. Using procedures provided in the API Data Book (Reference 13), the pseudocomponent composition of each product was used to estimate ASTM distillations, specific gravities and characterization factors. These in turn were used to estimate the product qualities such as viscosity, flash points, and pour points based on various published and in-house correlations. In particular, diesel cetane index was estimated using the ASTM calculations procedure (Reference 14).

The pseudocomponent approach was not used for estimating gasoline properties. Table 2-16 summarizes the RVP and RON used for each gasoline component. The RON values shown are in line with recently published figures (Reference 15). Octane blending, including the effect of lead additives was based on a method by Eastman (Reference 16). RVP's for each gasoline grade were obtained by blending on a molar basis.

Average regional lead additive concentrations in motor gasoline were derived from data sources provided by Environment Canada (Reference 6). Quarterly gasoline lead concentrations supplied by the refining industry were segregated into regions and a weighted average annual concentration derived. These figures were used to develop the 'Base Case' for the regional computer model. Table 2-19 reports the lead concentration used.

2.0 METHODOLOGY

2.3 ASSUMPTIONS AND LIMITATIONS

The major simplifying assumptions are discussed in this section. As a result of these assumptions, it should be emphasized that the value of the study lies in illustrating trends rather than providing concrete recommendations. Study cases should be evaluated in terms of deviations from the base case.

Composite Refinery

The composite refinery approach assumes that all the refineries in each region are operating identically at constant conditions throughout the year. In reality individual refineries may deviate significantly from the composite model in terms of processing schemes, process yields, feedstocks and product slates. The operation of a refinery may also change considerably as a result of changed feedstock or seasonal changes in product demand. The following assumptions related to the crude unit are of particular significance:

a) A minimum LSR end point of 140°F was assumed. This assumption is important since it sets the minimum amount of LSR that can be routed to the gasoline pool. The clear RON of the gasoline pool is very sensitive to the volume of LSR, which has a low RON (68 to 73). However, the contribution of LSR to gasoline in each region is very uncertain since it can vary markedly from refinery to refinery. The degree in which LSR production can be varied in each refinery is limited by such factors as cooling water temperature, crude tower operating pressure and offgas compressor capacity.

b) The model assumes only one light gas oil sidestream. Many refineries produce light and heavy gas oil streams which would enhance blending flexibility in the distillate pool.

Other Processes

Major assumptions have been made relating to the following processes.

Hydrocracking: The yields and product qualities are assumed to be identical to those of catalytic cracking. This approximation is more severe in the Atlantic region where hydrocracking constitutes 39.5% of total cracking capacity. The proportion of hydrocracking is significantly less in other regions (9.5% in Ontario, 17.8% in Quebec, 1% in the Western Region).

Coke, Asphalt, Aromatics and lube oil production: These processes are not represented. However, production as reported by Statistics Canada is relatively small. Nevertheless, in order to account for these products, allocations equal to the reported production rates were taken from the various streams as follows:-

Product	Stream
Asphalt	Vacuum Residuum
Lube Oil	Heavy Vacuum Gas Oil
Coke	Vacuum Residuum
Naphtha Specialties	Reformate

Visbreaking: This process was not included in the general model.

Distillate Desulphurization: The function of this process as represented in the model is to remove sulphur only. All the feed is assumed to be recovered with no generation of light ends or naphtha. Also the feed properties are assumed to be unaltered.

Product Specifications

The following assumptions were made relating to the product specifications.

a) Due to the lack of published production figures, different product grades of light fuel oil, diesel fuel and heavy fuel oil were not considered. The specifications given in Table 2-14 are representative of the several grades of these products.

b) Not all of the specifications stipulated by the Government of Canada for each product could be considered due to unavailability of adequate quality estimation correlations. However, some additional specifications taken from in-house refinery planning data were included in the list of specifications.

Catalytic Reforming Capacity

Certain assumptions were required to be made relating to the throughput/severity capacity of existing reformers, as discussed below.

Concerning the 1979 installed reformer capacity, it should be appreciated that most of the reformer units in Canada are old and of the semi-regenerative type. The required clear octane levels of the gasoline pools during the past ten years have risen steadily due to the gradual introduction of unleaded grades and the trend to reduce the lead additive usage in gasoline due to potential health hazards. Reformer severities in 1979 expressed in terms of RON generally ranged from 95 - 98 RON. Due to the trend of increasing proportions of unleaded grades in the total gasoline pools, the existing reformer units are being 'stretched' to operate at ever increasing severities. There are several methods to debottleneck and revamp a catalytic reforming unit. Some refiners have replaced the mono-metallic catalyst with expensive bi-metallic catalyst and in consequence have gained the advantage of increased catalyst life and improved reformate yields whilst retaining the high operating pressures. A few refiners have performed major revamps of existing units to convert them to the continuous regeneration type. These modifications allow the units to operate at severities of around 100 RON or higher. Operation of existing units at higher severities would induce both capacity and mechanical contraints on the systems and reduce catalyst life significantly.

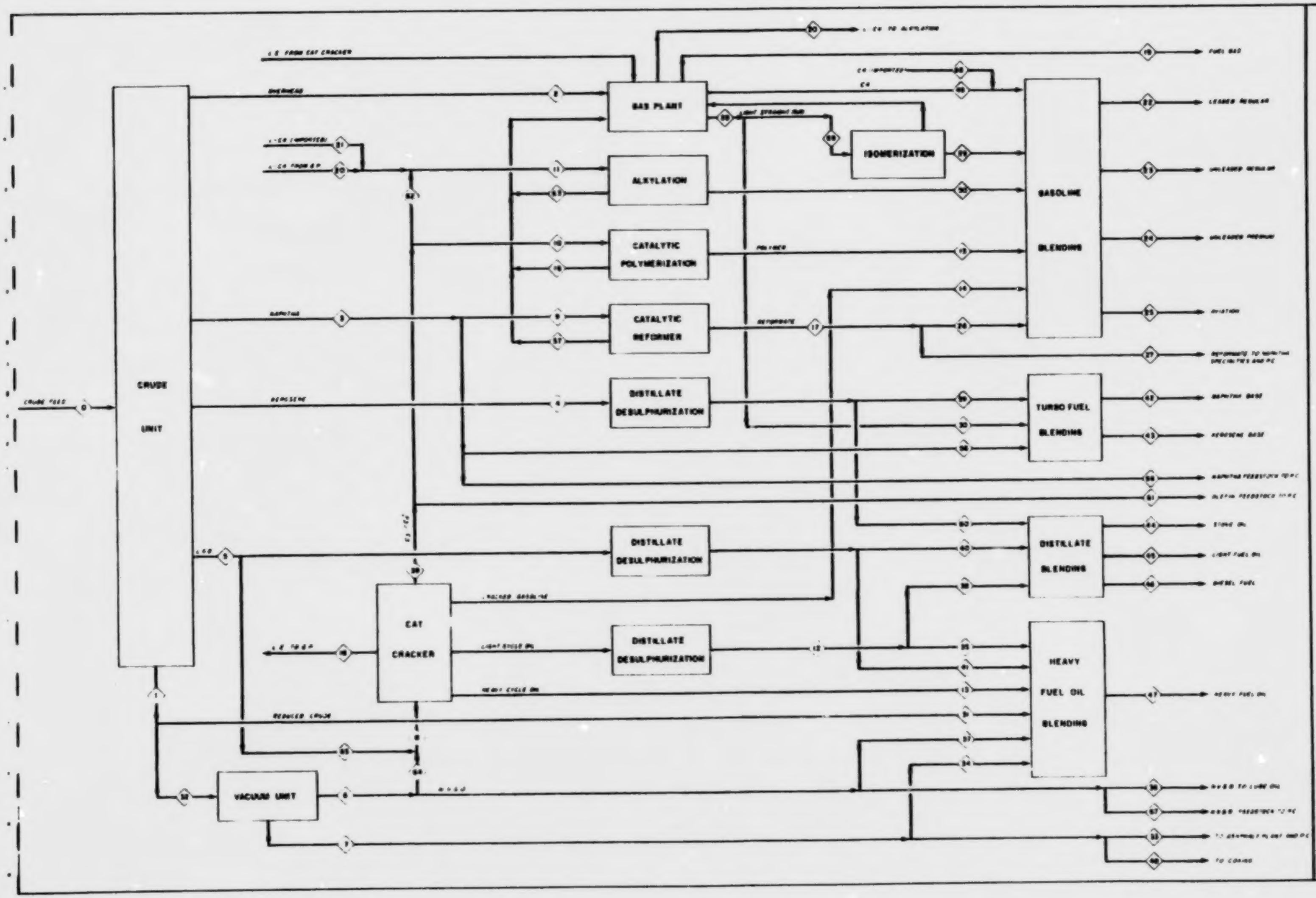
A hypothetical Quebec refinery study was conducted by ourselves in parallel to this one for Environment Canada. (Reference DSS File No: 4 DSS KE 204-0-0407) using the same tools; This study showed that advantages could be gained in terms of lower crude oil requirements, if the reforming severity was set at 103 RON, rather than 100 RON.

Consultations with UOP, a major U.S. licensor of installed Canadian catalytic reforming units, revealed that most units in Canada could not operate at severities of 103 RON for reasons explained previously. Since the exact status of reforming units throughout Canada could not be ascertained for this study, the initial approach adopted was one whereby the existing reformer capacity would be utilized to the maximum extent possible. Consequently, a realistic and practical maximum operating severity of 100 RON was adopted for the purposes of the study, but the installed capacity as reported in published literature would be referenced at 98 RON. Reformer process design/operating parameters were reviewed using in-house data and it was concluded that a 25% downrating factor should be applied to the 1979 installed reformer capacity, to allow continuous operation of the reformers at the 100 RON severity whilst retaining the same practical catalyst life factor.

Thus, if the installed capacity of a particular reforming unit was 20,000 BPCD in 1979, the equivalent capacity when operating at 100 RON would be only 15,000 BPCD.

If during the course of the computer iterations, it became manifest that all of the product specifications could not be attained with the reformer severity at 100 RON, then the severity was increased to

103 RON. For such cases, a conservative approach was adopted whereby all existing reformer units were declared redundant and new continuous type reformers were assumed to be needed for capital investment purposes. UOP consider that only four or five continuous type catalytic reforming units are installed in Canada to date.


CRUDE FEED
CRUDE UNIT
OVERHEAD
L.E. FROM CAT CRACKER
L-C4 (IMPORTED)
L C4 FROM G.P.
NAPHTHA
KEROSENE
L.G.O.
REDUCED CRUDE
VACUUM UNIT
H.V.G.O.
GAS PLANT
L C4 TO ALKYLATION
C4 IMPORTED
C4
LIGHT STRAIGHT RUN
ISOMERIZATION
ALKYLATION
CATALYTIC POLYMERIZATION
POLYMER
CATALYTIC REFORMER
REFORMATE
DISTILLATE DESULPHURIZATION
CAT CRACKER
L.E. TO G.P.
CRACKED GASOLINE
LIGHT CYCLE OIL
HEAVY CYCLE OIL
C3/C4
GASOLINE BLENDING
TURBO FUEL BLENDING
DISTILLATE BLENDING
HEAVY FUEL OIL BLENDING
FUEL GAS
LEADED REGULAR
UNLEADED REGULAR
UNLEADED PREMIUM
AVIATION
REFORMATE TO NAPHTHA SPECIALTIES AND PC
NAPHTHA BASE
KEROSENE BASE
NAPHTHA FEEDSTOCK TO PC
OLEFIN FEEDSTOCK TO PC
STOVE OIL
LIGHT FUEL OIL
DIESEL FUEL
HEAVY FUEL OIL
H.V.G.O. TO LUBE OIL
H.V.G.O. FEEDSTOCK TO PC
TO ASPHALT PLANT AND PC
TO COKING
0
1
2
3
4
5
6
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9
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63
64
65
67
LEGEND
H.V.G.O. HEAVY VACUUM GAS OIL
L.G.O. LIGHT GAS OIL
L.E. LIGHT ENDS (C4 AND LIGHTER)
G.P. GAS PLANT
F.G. FUEL GAS
P.C. PETROCHEMICAL PLANT
STREAM I.D. No. ON COMPUTER PRINTOUTS

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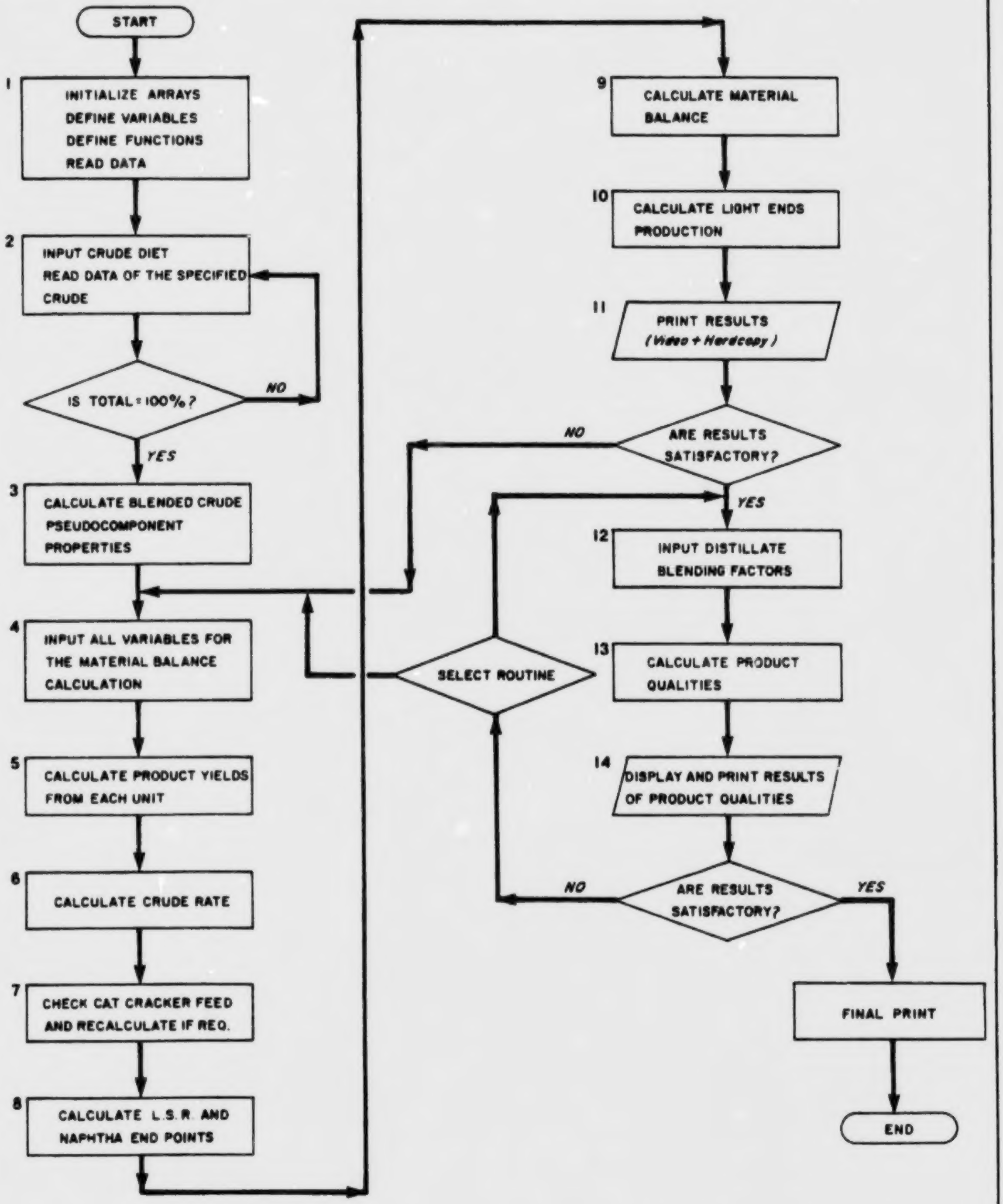
FIGURE 2-1
BLOCK FLOW
DIAGRAM OF
REFINERY MODEL


START
1
INITIALIZE ARRAYS
DEFINE VARIABLES
DEFINE FUNCTIONS
READ DATA
2
INPUT CRUDE DIET
READ DATA OF THE SPECIFIED CRUDE
IS TOTAL = 100%?
NO
YES
3
CALCULATE BLENDED CRUDE PSEUDOCOMPONENT PROPERTIES
4
INPUT ALL VARIABLES FOR THE MATERIAL BALANCE CALCULATION
5
CALCULATE PRODUCT YIELDS FROM EACH UNIT
6
CALCULATE CRUDE RATE
7
CHECK CAT CRACKER FEED AND RECALCULATE IF REQ.
8
CALCULATE L.S.R. AND NAPHTHA END POINTS
SELECT ROUTINE
9
CALCULATE MATERIAL BALANCE
10
CALCULATE LIGHT ENDS PRODUCTION
11
PRINT RESULTS (Video + Hardcopy)
ARE RESULTS SATISFACTORY?
NO
YES
12
INPUT DISTILLATE BLENDING FACTORS
13
CALCULATE PRODUCT QUALITIES
14
DISPLAY AND PRINT RESULTS OF PRODUCT QUALITIES
ARE RESULTS SATISFACTORY?
NO
YES
FINAL PRINT
END

FLOWCHART FOR REFINERY SIMULATION FIG. 2·2

WESTERN REGION
ONTARIO REGION
QUEBEC REGION
ATLANTIC REGION

DEFINITION OF REGIONS IN CANADA Figure 2·3

TABLE 2 1

ATLANTIC REGION
CRUDE DIET 1979

	LV %
VENEZUELAN LIGHT	31.0
VENEZUELAN HEAVY	3.5
ARABIAN LIGHT	51.7
ARABIAN HEAVY	3.5
IRANIAN HEAVY	10.3
TOTAL	100.0

AVERAGE BLEND

Base Case Throughput BPCD	313425
Gravity, °API	31.88
Sulphur, wt%	1.68

TABLE2-2

QUEBEC REGION
CRUDE DIET 1979

VENEZUELAN LIGHT	28.5
VENEZUELAN HEAVY	7.3
ARABIAN LIGHT	10.7
DOMESTIC LIGHT	40.6
CONDENSATE	2.8
SYNCRUDE	4.4
DOMESTIC HEAVY	5.7
TOTAL	100.0

AVERAGE BLEND

Base Case Throughput, BPCD	505062
Gravity, °API	33.51
Sulphur, wt.%	1.02

TABLE 2-3

ONTARIO REGION
CRUDE DIET 1979

	LV%
DOMESTIC LIGHT	80.6
CONDENSATE	6.5
SYNCRUDE	6.0
DOMESTIC HEAVY	6.9
TOTAL	100.0

AVERAGE BLEND

Base Case Throughput, BPCD	597796
Gravity,°API	38.81
Sulphur wt.%	0.335

TABLE 2-4

WESTERN REGION
CRUDE DIET 1979

	LV%
DOMESTIC LIGHT	81.1
CONDENSATE	5.5
SYNCRUDE	5.9
DOMESTIC HEAVY	7.5
TOTAL	100.0

AVERAGE BLEND

Base Case Throughput, BPCD	518641
Gravity, °API	38.49
Sulphur, wt.%	0.934

TABLE 2-5

ATLANTIC REGION
TOTAL PRODUCT SLATE FOR 1979
(BASE CASE)

PRODUCT	BPCD	VOL %
FUEL GAS (FOE) (float)	10556	3.37
LEADED REGULAR GASOLINE	50073	16.00
UNLEADED REGULAR GASOLINE	20630	6.60
UNLEADED PREMIUM GASOLINE	6619	2.12
AVIATION GASOLINE	0	0
NAPHTHA SPECIALTIES	0	0
OLEFIN PETROCHEMICAL FEED	2416	0.77
NAPHTHA PETROCHEMICAL FEED	0	0
JET FUEL B	1726	0.55
JET FUEL A	7224	2.31
STOVE OIL	8025	2.56
DIESEL FUEL	30225	9.66
LIGHT FUEL OIL (#2)	70698	22.59
HEAVY FUEL OIL (#6)	99604	31.83
LUBE OIL	0	0
ASPHALT	5125	1.64
COKE	0	0
TOTAL PRODUCT	312921	100
CRUDE OIL FEED	313425	100
IMPORTED i-C4	0	

TABLE 2-6
QUEBEC REGION
TOTAL PRODUCT SLATE FOR 1979
(BASE CASE)

PRODUCT	BPCD	VOL %
FUEL GAS (float)	20300	4.00
LEADED REGULAR GASOLINE	85908	16.95
UNLEADED REGULAR GASOLINE	40288	7.95
UNLEADED PREMIUM GASOLINE	32044	6.32
AVIATION GASOLINE	1693	0.33
NAPHTHA SPECIALTIES	4605	0.91
OLEFIN PETROCHEMICAL FEED	1870	0.37
NAPHTHA PETROCHEMICAL FEED	19500	3.85
JET FUEL B	5130	1.01
JET FUEL A	16469	3.25
STOVE OIL	8085	1.59
DIESEL FUEL	49259	9.72
LIGHT FUEL OIL (#2)	92430	18.23
HEAVY FUEL OIL (#6)	102480	20.22
LUBE OIL	4378	0.86
ASPHALT	22484	4.44
COKE	0	0
TOTAL PRODUCT	506923	100
CRUDE OIL FEED	505062	100
IMPORTED i-C4	0	

TABLE 2-7
ONTARIO REGION
TOTAL PRODUCT SLATE FOR 1979
(BASE CASE)

PRODUCT	BPCD	VOL %
FUEL GAS (float)	27454	4.50
LEADED REGULAR GASOLINE	135858	22.27
UNLEADED REGULAR GASOLINE	62423	10.23
UNLEADED PREMIUM GASOLINE	18213	2.99
AVIATION GASOLINE	221	0.04
NAPHTHA SPECIALTIES	11557	1.89
OLEFIN PETROCHEMICAL FEED	6200	1.02
NAPHTHA PETROCHEMICAL FEED	53830	8.82
JET FUEL B	3555	0.58
JET FUEL A	18313	3.00
STOVE OIL	4329	0.71
DIESEL FUEL	55060	9.03
LIGHT FUEL OIL (#2)	94756	15.54
HEAVY FUEL OIL (#6)	102698	16.84
LUBE OIL	7839	1.29
ASPHALT	6510	1.07
COKE	1095	0.18
TOTAL PRODUCT	609911	100
CRUDE OIL FEED	597796	99.35
IMPORTED i-C4	3895	0.65

TABLE 2-8
WESTERN REGION
TOTAL PRODUCT SLATE FOR 1979
(BASE CASE)

PRODUCT	BPCD	VOL %
FUEL GAS (float)	25120	4.80
LEADED REGULAR GASOLINE	159718	30.52
UNLEADED REGULAR GASOLINE	58307	11.14
UNLEADED PREMIUM GASOLINE	13350	2.55
AVIATION GASOLINE	2635	0.50
NAPHTHA SPECIALTIES	740	0.14
OLEFIN PETROCHEMICAL FEED	0	0
NAPHTHA PETROCHEMICAL FEED	800	0.15
JET FUEL B	18366	3.51
JET FUEL A	12686	2.42
STOVE OIL	11768	2.25
DIESEL FUEL	123877	23.67
LIGHT FUEL OIL (#2)	26869	5.14
HEAVY FUEL OIL (#6)	38536	7.36
LUBE OIL	3886	0.74
ASPHALT	24489	4.68
COKE	2250	0.43
TOTAL PRODUCT	523397	100
CRUDE OIL FEED	518641	99.77
IMPORTED i-C4	1218	0.23

TABLE 2-9

ATLANTIC REGION

INSTALLED REFINERY PROCESS UNIT CAPACITIES IN 1979

UNIT TYPE	BPCD
CRUDE DISTILLATION	448900
VACUUM DISTILLATION	110300
CATALYTIC REFORMING	63000
CATALYTIC CRACKING	83500
DISTILLATE DESULPHURIZATION	125700
POLYMERIZATION	2500
ALKYLATION	0
ISOMERIZATION	0

TABLE 2-10

QUEBEC REGION

INSTALLED REFINERY PROCESS UNIT CAPACITIES IN 1979

UNIT TYPE	BPCD
CRUDE DISTILLATION	617400
VACUUM DISTILLATION	233000
CATALYTIC REFORMING	112500
CATALYTIC CRACKING	164145
DISTILLATE DESULPHURIZATION	161200
POLYMERIZATION	6010
ALKYLATION	6100
ISOMERIZATION	0

TABLE 2-11

ONTARIO REGION

INSTALLED REFINERY PROCESS UNIT CAPACITIES IN 1979

UNIT TYPE	BPCD
CRUDE DISTILLATION	758200
VACUUM DISTILLATION	179200
CATALYTIC FEFORMING	140400
CATALYTIC CRACKING	172900
DISTILLATE DESULPHURIZATION	81100
POLYMERIZATION	2190
ALKYLATION	24670
ISOMERIZATION	0

TABLE 2-12

WESTERN REGION

INSTALLED REFINERY PROCESS UNIT CAPACITIES IN 1979

UNIT TYPE	BPCD
CRUDE DISTILLATION	535100
VACUUM DISTILLATION	173200
CATALYTIC REFORMING	93920
CATALYTIC CRACKING	152490
DISTILLATE DESULPHURIZATION	115460
POLYMERIZATION	5220
ALKYLATION	21070
ISOMERIZATION	0

TABLE 2-13

REFINERY OPERATING CONSTRAINTS

LSR end point:	140 - 210°F
Naphtha end point:	300 - 400°F
Kerosene end point:	480 - 550°F
Light gas oil end point:	600 - 700°F
Heavy Vacuum Gas Oil end point:	900 - 1050°F
Naphtha in Jet B	0.5 - 0.9
Naphtha in Jet A	0 - 0.25
LSR in Jet B	0 - 0.2
LCO end point	650°F (fixed)
Catalytic Cracker Conversion	50 - 80 wt %
Polymerization Turndown/Alkylation Turndown	.95 - 1.05
Reformer Severity	103 RON MAX

TABLE 2-14

GENERAL PRODUCT SPECIFICATIONS 1979		Base Case
Leaded Regular Gasoline		
Octane No.	RON	93.1
Lead Content	CC/IG	see regional concentration
RVP	PSIA	7.0 - 12.5
Unleaded Regular Gasoline		
Octane No.	RON	93.6
RVP	PSIA	7.0 - 12.5
Unleaded Premium Gasoline		
Octane No.	RON	97.5
RVP	PSIA	7.0 - 12.5
Aviation Gasoline		
Octane No.	RON	93.0
RVP	PSIA	10.0
Turbo Fuel Jet B (Naphtha Base)		
Specific Gravity		0.750-0.801
ASTM Distillation % 20/50/90, ^{o}F		290/370/470(max)
RVP	PSIA	2-3
Sulphur	Wt.%	0.4(max)
Turbo Fuel Jet A (Kerosene Base)		
Specific Gravity		0.774-.839
Pour Point ^{o}F		-50 (max)
ASTM Distillation 10%, ^{o}F		400 (max)

B

Flash Point °F	90 (min)
Sulphur wt %	0.2 (max)

Diesel Fuel

Pour Point °F	-25 (max)
Specific Gravity	.816 - 0.855
ASTM Distillation % 10/50/90, °F	460/520/625 (max)
Flash Point °F	130 (min)
Cetane Index	45 (min)
Viscosity @ 100°F, CS	1.85 - 3.95
Sulphur wt %	0.4 (max)

Stove Oil

Pour Point °F	-35 (max)
Specific Gravity	0.784 - 0.840
ASTM Distillation %, 10/50/90, °F	420/450/520 (max)
Flash Point °F	120 (min)
Viscosity @ 100°F CS	1.25 (min)
Sulphur wt %	0.2 (max)

Light Fuel Oil (#2)

Pour Point °F	0 (max)
Specific Gravity	0.820 - 0.887
ASTM Distillation % 10/50/90, °F	460/530/675 (max)
Flash Point °F	130 (min)
Viscosity @ 100°F, CS	1.75 - 4.25
Sulphur wt %	0.5 (max)

C

Heavy Fuel Oil

Sulphur	wt %	3.5 (max)
Viscosity @ 122°F,	CS	92 - 638

TABLE 2-15

SUMMARY OF CRUDE ANALYSES

Source/Type	Gravity °API	WT% S	Specific Assays Used
Domestic			
Light	39.0	.24/1.15	IPL Blend #1/#2
Heavy	25.1	2.39	Bow Bell
Synthetic	27.7	.08	Syncrude Synthetic
Condensate	63.0	.095	Rimbey
Imported			
Venezuelan Light	30.4	1.36	Ceuta
Venezuelan Heavy	16.8	2.40	Bachaquero
Arabian Light	34.4	1.70	Arabian Light
Arabian Heavy	28.2	2.84	Arabian Heavy
Iranian Heavy	30.8	1.60	Iranian Heavy

TABLE 2-16

GASOLINE COMPONENT PROPERTIES

	RVP	RON	Sensitivity (RON-MON)	WT% S
Reformate	5.0	a	0.9(RON)-1	0
Alkylate	8.5	93.	2.2	0
Polymer	4.5	96.5	14.0	0
Crackate	4.5	b	0.925(RON)-4.25	f
Light Straight Run (LSR)	c	c	c	c
Isomerate	e	d	e	e
C_4	55.	93.	1.0	0

Notes:
a. Specified for each case
b. Varies between 89 and 92.2 depending on conversion
c. Estimated from crude assay data
d. RON of LSR +13.0
e. Same as for LSR
f. Estimated from cat feed sulphur content

TABLE 2-17
REGIONAL FEEDSTOCK, PRODUCT AND UTILITY PRICES FOR 1979

ITEM	UNIT	ATLANTIC	QUEBEC	ONTARIO	WESTERN	SOURCE OF INFORMATION
ELECTRICITY	¢/KWh	3.16	1.80	1.95	1.50	ONTARIO HYDRO & MECO
CRUDE PRICE	$/BBL	23.18	18.88	14.73	13.75	EMR & CORPUS
HFO	$/BBL	17.35	18.83	19.83	18.04	EMR & CORPUS (ENERGY ANALECT MARCH 1979)
NO. OF REFINERIES		5	7	8	17	
NATURAL GAS	$/1000 scf	----	2.53	2.30	1.56	STATS CAN
i-C4	$/BBL	39.96	39.96	39.96	39.96	CHEMICAL ENG. MAGAZINE
LEAD ADDITIVES	¢/lb	91.4	91.4	91.4	91.4	ETHYL CANADA

NOTE: Natural Gas prices quoted above were not used in the study calculations, but are reported here for information only.

. MMT is approximately twice as costly as TEL for the same octane number boost.

TABLE 2-18

PROCESS UNIT ENERGY REQUIREMENT

PROCESS UNIT	POWER 10^3 BTU/bbl	FUEL 10^3 BTU/bbl	STEAM 10^3 BTU/bbl	TOTAL 10^3 BTU/bbl Feed
Crude	6	80	9	95
Vacuum	2	45	13	60
Cracker	60	79	-39	100
Poly	28	200	272	500
Alkyl	35	450	15	500
Reformer	30	294	26	350 @ 98 RON 365 @ 100 RON 380 @ 103 RON
Naphtha Hydrotreater	14	66	10	90
Isomerization	15	40	45	100
Gas Plant	25	--	575	600
Desalter	1	--	–	1

TABLE 2-19

1979 LEAD CONSUMPTION BY CANADIAN REFINERIES FOR PRODUCTION OF LEADED REGULAR GASOLINE.

REGION	FIRST QUARTER		SECOND QUARTER		THIRD QUARTER		FOURTH QUARTER		TOTAL	
	VOL MM IG	WEIGTED AVERAGE LEAD CONCENTRATION GM/IG	VOL MM IG	AVERAGE GM/IG	VOL MM IG	AVERAGE GM/IG	VOL MM IG	AVERAGE GM/IG	VOL MM IG	AVERAGE GM/IG
ATLANTIC	142.61	2.27	133.64	2.78	146.94	2.97	140.96	2.80	564.15	2.71
QUEBEC	267.21	2.41	257.18	3.00	268.51	2.92	246.23	2.69	1039.13	2.76
ONTARIO	392.06	1.61	410.13	1.80	354.42	2.01	428.34	1.50	1584.95	1.73
WESTERN	510.95	1.53	560.21	2.09	634.91	2.63	622.12	1.72	2328.19	1.99

TABLE 2-20

REGIONAL CRUDE OIL YIELD STRUCTURES - 1979

STREAM	TBP CUT WIDTH OF	LV % REGION			
		ATLANTIC	QUEBEC	ONTARIO	WESTERN
GAS		2.0	2.9	4.1	4.1
LSR	60-140	3.6	4.3	6.3	6.0
NAPHTHA	140-340	16.9	20.0	25.2	24.8
KEROSENE	340-520	17.9	17.2	17.9	18.0
LIGHT GAS OIL	520-660	13.1	13.1	13.1	13.2
HEAVY VACUUM GAS OIL	660-1000	26.8	25.6	22.0	22.3
VACUUM RESIDUUM	1000+	19.7	16.9	11.4	11.6

TABLE 2 - 21

1979 MOTOR GASOLINE POOL GRADE DISTRIBUTION BY REGION

Region	Vol % Gasoline Grade		
	Leaded Regular	Unleaded Regular	Unleaded Premium
Atlantic	64.8	26.6	8.6
Quebec	54.3	25.5	20.2
Ontario	62.7	28.9	8.4
Western	69.0	25.2	5.8

TABLE 2-22

SUMMARY OF MOTOR GASOLINE OCTANE QUALITY FOR SIMULATIONS

REGION -	CASE	Octane Number RON		
		Leaded Regular	Unleaded Regular	Unleaded Premium
ATLANTIC	BASE	93.1	93.6	97.5
	2 LEAD	93.6	93.6	97.5
	1 LEAD	93.6	93.6	97.5
	0 LEAD	93.1	93.6	96.5
QUEBEC	BASE	93.1	93.6	97.5
	2 LEAD	93.6	93.6	97.5
	1 LEAD	93.6	93.6	97.5
	0 LEAD	92.9	92.9	96.5
ONTARIO	BASE	93.1	93.6	97.5
	1 LEAD	93.6	93.6	97.5
	0 LEAD	93.1	93.6	96.5
WESTERN	BASE	93.1	93.6	97.5
	1 LEAD	93.6	93.6	97.5
	0 LEAD	93.6	93.6	97.5

SECTION 3

3.0 DISCUSSION OF CASE SIMULATIONS

3.1 BASE CASE

The objectives for developing the base cases were as follows:-

1. Determine whether the model adequately represents refinery industry in each region.

2. Provide a basis for identifying trends and estimating energy penalty and operating costs resulting from the different lead levels.

As discussed in 2.1 the model was essentially tested using the following 1979 published data:

Crude diets - Tables 2-1 to 2-4
Product slates - Tables 2-5 to 2-8
Unit capacities - Tables 2-9 to 2-12
Lead and octane levels - Table 2-19 and Reference 2

Crude rate was minimized in order to provide a consistent basis for evaluating capital costs and energy penalties.

A reformer severity of 98 RON was required to satisfy the premium gasoline RON of 97.5.

In general, catalytic reformer throughputs used in the base case were very low (47.4% in Ontario to 70.8% in Western region) while cat cracker throughputs were high (100% in Ontario, Western and Quebec). All polymerization and alkylation units were at 100%. Cat cracker, alkylation and polymerization throughput was favoured at the expense of catalytic reformer throughput because this provided minimum crude rates.

All crude rates estimated in the base cases are slightly lower than published figures (Reference 7) as shown below:

	BPCD		
	Base Case	Reported	% difference
Atlantic	313425	321503	2.6
Quebec	505062	509878	.96
Ontario	597796	600054	.38
Western	518644	528502	1.9

The discrepancies are probably within the accuracy of the reported data. The base case values tend to be lower due to the fact that volume losses resulting from desulphurization, visbreaking, deasphalting, and coking were not accounted for.

The distillate product qualities for all regions were well within the specifications.

3.0 DISCUSSION OF CASE SIMULATIONS

3.2 DEVELOPMENT OF LOW-LEAD CASES

Due to time limitations, the investments considered in this study were confined to upgrading naphtha (by additional catalytic reforming) and LSR (by isomerization) in order to achieve the required RON levels. Increased alkylation capacity was another parameter considered for the zero lead cases only to reduce net LPG production. Other options such as increasing RON of crackers or increasing production of polymerization were not evaluated but are discussed qualitatively in Section 5.2.1.

Six principal methods of improving octane were applied to develop the study cases:

1. Increase in reformer throughput. In order to provide additional reformer feed, the naphtha boiling range must be extended by a) lowering the initial boiling point or b) increasing the end point.

   a) Lowering the initial boiling point was found to be the desirable route in spite of lower liquid yields because it resulted in displacing LSR (68-73 RON) from the gasoline pool with high octane reformate. The extent to which this end point could be lowered was limited by the LSR end point limitation of 140°F.

   b) Increasing the naphtha end point is less desirable because it results in displacing FCC gasolines (crackate, polymer, and alkylates) from the gasoline pool thereby resulting in a smaller increase in the overall octane of the gasoline pool per additional barrel of reformate. In addition, increasing the naphtha end point reduces the kerosene fraction thereby adversely affecting distillate product quality.

2a. Increase reformer severity. This measure allows an increase in the octane-barrel contribution of reformate to the gasoline pool without infringing on kerosene availability for the distillate pool. As pointed out above, a severity of 98 RON was required for the base case. For the study cases, it was assumed that severity could be increased by only 2 additional octane points due to limitations of existing units (see section 2.3). In addition, it was assumed that the reformer capacity was reduced to 75% at this higher RON level.

2b. When all product specifications could not be achieved, severity was raised to 103 RON, and existing reformer capacity was considered redundant.

3. Addition of isomerization. An isomerization unit with no recycle yielding an increase of 13 RON points was introduced as required. This unit was particularly useful in cases where the 140°F LSR minimum end point limitation was reached such that additional increases in reformer feed would adversely affect distillate product quality. Application of this measure automatically implies capital investment since there are no known isomerization units presently operating in Canada.

4. Increase FCC gasoline make. This measure raises the gasoline pool RON provided that the additional FCC gasoline displaces LSR. This approach is similar to 1.a above in which reformate is used to displace LSR. In general, increasing FCC gasoline make is preferable since it lowers crude rates and improves distillate qualities.

5. Increases in FCC conversion. This measure provides marginal increase in RON for a constant FCC gasoline make as a result of the higher severity (see Table 2-16). The higher severity also increases available olefin feedstock.

6. Increase in alkylation capacity. Once the above-mentioned parameters had been exploited, a 'fine tuning' exercise was executed by increasing alkylation capacity together with increased cracker conversion, to ascertain the potential for reducing crude oil consumption. Simultaneously this would have the effect of reducing net LPG production through consumption of surplus i-C_4.

The specific measures used in the study cases were as follows:-

Case	2 Lead	1 Lead	NO. Lead
Atlantic	1a, 2	1a, 1b, 2	1a, 1b, 2^b, 3, 6
Quebec	1a, 1b, 2, 5	1a, 1b, 2, 5	1a, 1b, 2^b, 3, 6
Ontario	-	1a, 1b, 2, 5	1a, 1b, 2^b, 3, 6
Western	-	1a	1a, 2^a, 3, 4, 5

As discussed in Section 2.1, the selection for each case was based on a trial and error search geared toward minimizing crude rate and minimizing isomerization and additional reformer capacity requirements.

Highlights are as follows:-

1. Isomerization units were introduced in all regions (#3) for the no-lead cases only.

2. Reformer throughput was increased (1a and 1b) in all regions for all cases. However, throughput exceeded installed capacity only

in the no-lead case. Consequently, capital investment was found to be required for the no-lead cases only.

3. Distillate product specifications could be met in all regions without too much difficulty. The situation was easier for the Western region because the increase in reformer throughput was taken at the expense of LSR (1a) rather than kerosene thereby eliminating any adverse effect on distillate properties.

4. Reformer severity was increased for all cases except Western (1 lead); Additionally, severities of 103 RON were required for all regions except Western, which could remain at 100 RON, for all the zero lead cases.

5. Alkylation capacity increases were effective for all regions except Western, in reducing crude oil consumption in the zero lead cases.

3.0 DISCUSSION OF CASE SIMULATIONS

3.3 CAPITAL COST REQUIREMENTS

As pointed out in the previous section, capital costs were required for the no-lead cases only. The extent of additional capacity requirements, estimated in this study is dependent on the three following major factors related to the base case:

1. Additional octane-barrel requirements to convert all of the leaded regular to unleaded regular. This value depends on the clear RON estimated in the base case as well as the volume of the leaded regular that is produced.

2. Excess reformer capacity.

3. Margin between the LSR end point calculated in the base case and the minimum LSR end point limitation of 140°F imposed in this study. This margin is a measure of how much light straight run could be displaced from the gasoline pool by high octane components.

These parameters are summarized in Table 3-1 along with the isomerization and additional reformer capacity requirements. The total additional capacity requirement is not highest in the West in spite of the high reformer utilization in the base case as well as the high octane-barrel requirement. This is accounted for by the fact that a large volume of LSR was displaced from the gasoline pool with cracked gasoline and reformate by measures 1a, 4 and 5 discussed above. This is reflected by a 40°F drop in the LSR end point (191°F to 151°F).

TABLE 3 - 1

BASIS FOR ADDITIONAL CAPACITY REQUIREMENTS FOR NO LEAD CASE

BASE CASE	ATLANTIC	QUEBEC	ONTARIO	WESTERN	CANADA
Leaded Regular, BPCD	50,073	85,908	135,858	159,718	431,557
RON (clear)	86.2	85.7	87.0	86.4	86.4
LSR end-point, °F	159	142	143	191	
Reformer Utilization, %	57.8	54.9	47.4	70.8	
NO-LEAD CASE					
Octane-BBL requirement, BPCD-RON	338,885	558,292	810,521	1,115,000	2,822,698
LSR end-point, °F	147	140	140	151	
Additional Reformer Capacity, BPCD	38,000	69,600	75,500	23,560	206,660
Isomerization Capacity, BPCD	12,837	21,853	38,320	36,938	109,948
Additional Alkylation Capacity, BPCD	1,746	1,387	4,216	-	7,349

3.0 DISCUSSIONS OF CASE SIMULATIONS

3.4 ENERGY PENALTIES

As expected, crude rates increase as lead level is reduced (see Tables 3-2 to 3-6). This is mainly a consequence of higher volume losses through the catalytic reformers, but these losses are recovered in the most part as LPG (or fuel gas).
Table 3-7 summarises the energy penalties for all the regional cases. The anomalies apparent in the Atlantic and Quebec zero lead cases are accounted for primarily by the masking effect of MMT usage.

It is interesting to note the need for i-C_4 imports to satisfy alkylation requirements. No i-C_4 is required in the Atlantic and Quebec regions for any case. For Western, i-C_4 imports reduce to zero as the lead level reduces to zero. However, for the Ontario region, imports of i-C_4 will be needed for all the cases, and is higher for the zero lead case compared to the base case.

Generally, the energy penalty will tend to increase with increasing lead level control for the same octane levels due to higher operating energy requirements and higher crude oil throughputs.
The estimated additional energy requirements for the process units are given in Tables 3-8 to 3-17

A typical calculation for the evaluation of energy penalty is given in Table 3-18

TABLE 3-2

SUMMARY OF RESULTS

ATLANTIC REGION

CASE	REFORMER SEVERITY RON	ISOMERIZATION UNIT INCLUSION	CRUDE OIL RATE BPCD	CRUDE THROUGHPUT INCREMENT %	HEAVY FUEL OIL PRODUCTION INCREASE %	FUEL GAS BPCD(FOE)	I-C4 IMPORTED BPCD	ENERGY PENALTY %
BASE	98	NO	313425	0	0	10556.4	-	-
2 LEAD	100	NO	314896	0.47	0	11907.8	-	0.13
1 LEAD	100	NO	316177	0.88	0	12988.2	-	0.35
NO LEAD	103	YES	316710	1.05	0	14371.7	-	0.15

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TABLE 3-3

SUMMARY OF RESULTS

QUEBEC REGION

CASE	REFORMER SEVERITY RON	ISOMERIZATION UNIT INCLUSION	CRUDE OIL RATE BPCD	CRUDE THROUGHPUT INCREMENT %	HEAVY FUEL OIL PRODUCTION INCREASE %	FUEL GAS BPCD(FOE)	I-C4 IMPORTED BPCD	ENERGY PENALTY %
BASE	98	NO	505062	-	-	20299.8	-	-
2 LEAD	100	NO	508495	0.68	-	23243.9	-	0.21
1 LEAD	100	NO	514849	1.94	-	26671.5	-	0.89
NO LEAD	103	YES	512700	1.51	-	27756.1	-	0.40

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TABLE 3-4

SUMMARY OF RESULTS

ONTARIO REGION

CASE	REFORMER SEVERITY RON	ISOMERIZATION UNIT INCLUSION	CRUDE OIL RATE BPCD	CRUDE THROUGHPUT INCREMENT %	HEAVY FUEL OIL PRODUCTION INCREASE %	FUEL GAS BPCD(FOE)	I-C4 IMPORTED BPCD	ENERGY PENALTY %
BASE	98	NO	597796	0	0	27453.6	3894.48	-
2 LEAD	-	-	-	-	-	-	-	-
1 LEAD	100	NO	605453	1.28	0	33407.9	3416.39	0.37
NO LEAD	103	YES	607404	1.61	0	38062.2	5209.52	0.46

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TABLE 3-5

SUMMARY OF RESULTS

WESTERN REGION

CASE	REFORMER SEVERITY RON	ISOMERIZATION UNIT INCLUSION	CRUDE OIL RATE BPCD	CRUDE THROUGHPUT INCREMENT %	HEAVY FUEL OIL PRODUCTION INCREASE %	FUEL GAS BPCD(FOE)	I-C4 IMPORTED BPCD	ENERGY PENALTY %
BASE	98	NO	518641	-	-	25120.3	1218.43	-
2 LEAD	-	-	-	-	-	-	-	-
1 LEAD	98	NO	521864	0.62	-	27799.3	648.65	0.38
NO LEAD	100	YES	530382	2.26	-	35479.2	-	0.85

TABLE 3-6

SUMMARY OF RESULTS

CANADA

REGION	CASE	REFORMER SEVERITY RON	CRUDE OIL RATE BPCD	CRUDE THROUGHPUT INCREMENT %	FUEL GAS PRODUCTION BPCD (FOE)	i-C4 IMPORTED BPCD	ENERGY PENALTY %
ATLANTIC	BASE	98	313425	-	10556.4	-	-
	2 BASE	100	314896	0.47	11907.8	-	0.13
	1 LEAD	100	316177	0.88	12988.2	-	0.35
	0 LEAD	103	316710	1.05	14371.7	-	0.15
QUEBEC	BASE	98	505062	-	20299.8	-	-
	2 LEAD	100	508495	0.68	23243.9	-	0.21
	1 LEAD	100	514849	1.94	26671.5	-	0.89
	0 LEAD	103	512700	1.51	27756.1	-	0.40
ONTARIO	BASE	98	597796	-	27453.6	3894.48	-
	1 LEAD	100	605453	1.28	33407.9	3416.39	0.37
	0 LEAD	103	607404	1.61	38062.2	5209.52	0.46
WESTERN	BASE	98	518641	-	25120.3	1218.43	-
	1 LEAD	98	521864	0.62	27799.3	648.65	0.38
	0 LEAD	100	530382	2.26	35479.2	-	0.85
CANADA	BASE		1934924	-	83430.1		
	2 LEAD		1939828	0.25	87725.6		
	1 LEAD		1958343	1.21	100866.9		
	0 LEAD		1967196	1.67	115669.2		

TABLE 3-7

SUMMARY OF ENERGY PENALTY BREAKDOWN

% ENERGY PENALTY

CONTROL CASE

	2 LEAD	1 LEAD	NO LEAD
ATLANTIC			
CRUDE	0.47	0.88	1.04
i-C4	0	0	0
FUEL GAS	-0.47	-0.85	-1.32
HFO	0	0	0
PROCESS ENERGY	0.13	0.32	0.43
TOTAL	0.13	0.35	0.15
QUEBEC			
CRUDE	0.68	1.94	1.51
i-C4	0	0	0
FUEL GAS	-0.64	-1.39	-1.62
HFO	0	0	0
PROCESS ENERGY	0.17	0.34	0.51
TOTAL	0.21	0.89	0.40
ONTARIO			
CRUDE		1.28	1.61
i-C4		-0.06	0.16
FUEL GAS		-1.12	-1.99
HFO		0	0
PROCESS ENERGY		0.27	0.68
TOTAL	0	0.37	0.46
WESTERN			
CRUDE		0.62	2.26
i-C4		-0.08	-0.17
FUEL GAS		-0.58	-2.24
HFO		0	0
PROCESS ENERGY		0.42	1.0
TOTAL	0	0.38	0.85

SIMULATION RESULTS TABLE 3-8

ATLANTIC CASE: 2 LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^6BTU/CD)
CRUDE DISTILLATION	314896	313425	1471	95	139.75
VACUUM DISTILLATION	102204	101597	607	60	36.42
CATALYTIC CRACKER	63000	63000	-	100	-
CATALYTIC POLYMERIZATION	6046.8	6044.5	2.3	500	1.15
HF ALKYLATION	-	-	-	500	-
CATALYTIC REFORMER	38300	36400	1900	365	693.50
HYDROTREATER NAPHTHA	38300	36400	1900	90	171.00
ISOMERIZATION	-	-	-	100	-
HDS-LSR	-	-	-	90	-
GAS PLANT	22698.5	21250	1448.5	600	869.10
DESALTING	314896	313425	1471	1	1.47
FUEL GAS (FOE)	11907.8	10556.4	1351.4	TOTAL	1912.39
HFO PRODUCTION	99604	99604	-		
IMPORTED i-C4	-	-	-		

SIMULATION RESULTS TABLE 3-9

ATLANTIC CASE: 1 LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^6BTU/CD)
CRUDE DISTILLATION	316177	313425	2752	95	261.44
VACUUM DISTILLATION	109597	101597	8000	60	480.00
CATALYTIC CRACKER	63000	63000	-	100	-
CATALYTIC POLYMERIZATION	6048.0	6044.5	3.5	500	1.75
HF ALKYLATION	-	-	-	500	-
CATALYTIC REFORMER	42900	36400	6500	365	2372.50
HYDROTREATER NAPHTHA	42900	36400	6500	90	585.00
ISOMERIZATION	-	-	-	100	-
HDS-LSR	-	-	-	90	-
GAS PLANT	23976.3	21250	2726.3	600	1635.78
DESALTING	316177	313425	2752	1	2.75
				TOTAL	5339.22
FUEL GAS (FOE)	12988.2	10556.4	2431.8		
HFO PRODUCTION	99604	99604	-		
IMPORTED i-C4	-	-	-		

SUMMARY OF SIMULATION RESULTS TABLE 3-10

ATLANTIC CASE NO LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^6 BTU/CD
CRUDE DISTILLATION	316,710	313,425	3,285	95	312.08
VACUUM DISTILLATION	81,555	101,597	20,042	60	- 1202.52
CATALYTIC CRACKER	64,000	63,000	1,000	100	100.00
CATALYTIC POLYMERIZATION	6,050	6,044.5	5.5	500	2.75
HF ALKYLATION	2,423	-	2,423	500	1,211.50
CATALYTIC REFORMER	38,000	36,400	1,600	380	608.00
HYDROTREATER NAPHTHA	38,000	36,400	1,600	90	144.00
ISOMERIZATION	12,836.7	-	12,836.7	100	1,283.67
HDS-LSR	12,836.7	-	12,836.7	90	1,155.30
GAS PLANT	26,507.3	21,250	5,257.3	600	3,154.38
DESALTING	316,710	313,425	3,285	1	3.29
FUEL GAS (FOE)	14,371.7	10,556.4	3,815.3	TOTAL	6,772.45
HFO PRODUCTION	99,604.0	99,604.0	-		
IMPORTED i - C4					

SIMULATION RESULTS

TABLE 3-11

QUEBEC CASE: 2 LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^{6}BTU/CD)
CRUDE DISTILLATION	508495	505062	3433	95	326.14
VACUUM DISTILLATION	232557	233321	-764	60	-45.84
CATALYTIC CRACKER	155078	162281	-7203	100	-720.30
CATALYTIC POLYMERIZATION	14500	14580	-80	500	-40.00
HF ALKYLATION	8295.4	8440	-144.6	500	-72.30
CATALYTIC REFORMER	68000	61800	6200	365	2263.00
HYDROTREATER NAPHTHA	68000	61800	6200	90	558.00
ISOMERIZATION	-	-	-	100	-
HDS-LSR	-	-	-	90	-
GAS PLANT	49426.7	46516	2910.7	600	1746.42
DESALTING	508495	505062	3433	1	3.43
FUEL GAS (FOE)	23243.9	20299.8	2944.1	TOTAL	4018.55
HFO PRODUCTION	102480	102480	-		
IMPORTED i-C4	-	-	-		

SIMULATION RESULTS TABLE 3-12

QUEBEC CASE: 1 LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^6BTU/CD)
CRUDE DISTILLATION	514849	505062	9787	95	929.77
VACUUM DISTILLATION	223659	233321	-9662	60	-579.72
CATALYTIC CRACKER	122377	162281	-39904	100	-3990.40
CATALYTIC POLYMERIZATION	13300	14580	-1280	500	-640.00
HF ALKYLATION	7738.7	8440	-701.3	500	-350.65
CATALYTIC REFORMER	84300	61800	22500	365	8212.50
HYDROTREATER NAPHTHA	84300	61800	22500	90	2025.00
ISOMERIZATION	-	-	-	100	-
HDS-LSR	-	-	-	90	-
GAS PLANT	52224.6	46516	5708.6	600	3425.16
DESALTING	514849	505062	9787	1	9.79
				TOTAL	9041.45
FUEL GAS (FOE)	26671.5	20299.8	6371.7		
HFO PRODUCTION	102480	102480	-		
IMPORTED i-C4	-	-	-		

SUMMARY OF SIMULATION RESULTS TABLE 3 - 13

QUEBEC CASE: NO LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^6 BTU/CD)
CRUDE DISTILLATION	512,700	505,062	7,638	95	725.61
VACUUM DISTILLATION	227,215	233,321	-6,106	60	- 366.36
CATALYTIC CRACKER	152,178	162,281	-10,103	100	-1,010.30
CATALYTIC POLYMERIZATION	14,300	14,580	-280	500	-140.00
HF ALKYLATION	10,383	8,440	1,943	500	971.50
CATALYTIC REFORMER	69,600	61,800	7,800	380	2,964.00
HYDROTREATER NAPHTHA	69,600	61,800	7,800	90	702.00
ISOMERIZATION	21,853	-	21,853	100	2,185.30
HDS-LSR	21,853	-	21,853	90	1,966.77
GAS PLANT	54,951	46,516	8,435	600	5,061.00
DESALTING	512,700	505,062	7,638	1	7.64
FUEL GAS (FOE)	27,756.1	20,299.8	7,456.3	TOTAL	13,067.16
HFO PRODUCTION	102,480	102,480	-		
IMPORTED i - C4					

SIMULATION RESULTS TABLE 3-14

ONTARIO CASE: 1 LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^6BTU/CD)
CRUDE DISTILLATION	605453	597796	7657	95	727.42
VACUUM DISTILLATION	147991	159331	-11340	60	- 680.40
CATALYTIC CRACKER	141100	172900	-31800	100	-3180.00
CATALYTIC POLYMERIZATION	53 50	5350	-	500	-
HF ALKYLATION	3410.3	34150	-48.7	500	-24.35
CATALYTIC REFORMER	83000	66500	16500	365	6022.50
HYDROTREATER NAPHTHA	83000	66500	16500	90	1485.00
ISOMERIZATION	-	-	-	100	-
HDS-LSR	-	-	-	90	-
GAS PLANT	70828.8	64432.5	6396.3	600	3837.78
DESALTING	605453	597796	7657	1	7.66
				TOTAL	8195.61
FUEL GAS (FOE)	33407.9	27453.6	5954.3		
HFO PRODUCTION	102698	102698.0	-		
IMPORTED i-C4	3416.39	3894.48	-478.09		

SUMMARY OF SIMULATION RESULTS TABLE 3-15

ONTARIO CASE NO LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^6 BTU/CD)
CRUDE DISTILLATION	607,404	597,796	9,608	95	912.76
VACUUM DISTILLATION	171,253	159,331	11,922	60	715.32
CATALYTIC CRACKER	147,600	172,900	-25,300	100	- 2,530.00
CATALYTIC POLYMERIZATION	4,600	5,350	- 750	500	- 375.00
HF ALKYLATION	40,090	34,150	5,940	500	2,970.00
CATALYTIC REFORMER	75,500	66,500	9,000	380	3,420.00
HYDROTREATER NAPHTHA	75,500	66,500	9,000	90	810.00
ISOMERIZATION	38,319.6	-	38,319.6	100	3,831.96
HDS-LSR	38,319.6	-	38,319.6	90	3,448.76
GAS PLANT	77,468.7	64,432.5	13,036.2	600	7,821.72
DESALTING	607,404	597,796	9,608	1	9.61
FUEL GAS (FOE)	38062.2	27,453.6	10,608.6	TOTAL	21,035.13
HFO PRODUCTION	102,698	102,698.0	-		
IMPORTED i - C4	5,209.52	3,894.48	1,315.04		

SIMULATION RESULTS TABLE 3-16

WESTERN CASE: 1 LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^6BTU/CD)
CRUDE DISTILLATION	521864	518641	3223	95	306.19
VACUUM DISTILLATION	163125	162118	1007	60	60.42
CATALYTIC CRACKER	151793	151891	-98	100	-9.80
CATALYTIC POLYMERIZATION	12700	12700	-	500	-
HF ALKYLATION	29275	29250	25	500	12.50
CATALYTIC REFORMER	88000	66500	21500	350	7525.00
HYDROTREATER NAPHTHA	88000	66500	21500	90	1935.00
ISOMERIZATION	-	-	-	100	-
HDS-LSR	-	-	-	-	-
GAS PLANT	66333.6	62264	4069.5	600	2441.70
DESALTING	521864	518641	32 23	1	3.22
FUEL GAS (FOE)	27799.3	25120.3	2679	TOTAL	12274.23
HFO PRODUCTION	38536	38536	-		
IMPORTED i-C4	648.65	1218.43	-569.78		

SIMULATION RESULTS TABLE 3-17

WESTERN CASE: NO LEAD

PROCESS UNIT	CASE (BPCD)	BASE (BPCD)	INCREMENT (BPCD)	UNIT ENERGY REQUIREMENT (10^3 BTU/bbl)	ADDITIONAL ENERGY REQUIREMENT (10^6BTU/CD)
CRUDE DISTILLATION	530382	518641	11741	95	1115.40
VACUUM DISTILLATION	172479	162118	10361	60	621.66
CATALYTIC CRACKER	150271	151891	-1620	100	-162.00
CATALYTIC POLYMERIZATION	12700	12700	-	500	-
HF ALKYLATION	29300	29250	50	500	25.00
CATALYTIC REFORMER	94000	66500	27500	365	10037.50
HYDROTREATER NAPHTHA	94000	66500	27500	90	2475.00
ISOMERIZATION	36938	-	36938	100	3693.80
HDS-LSR	36938	-	36938	90	3324.42
GAS PLANT	74412.6	62264	12148.6	600	7289.16
DESALTING	530382	518641	11741	1	11.74
				TOTAL	28431.68
FUEL GAS (FOE)	35479.2	25120.3	10358.9		
HFO PRODUCTION	38536	38536	-		
IMPORTED i-C4	-	1218.43	-1218.43		

TABLE 3 - 18

TYPICAL CALCULATION OF ENERGY PENALTY

ONTARIO

HEATING VALUE OF CRUDE = (18,320+40(138.81-10))x(10x0.8308)=161,777

ADDITIONAL CRUDE PROCESSED = 9,608 BPCD
= 9,608 BPCD x 161,777 $\frac{\text{BTU}}{\text{IG}}$ x 35 IG/BBL
= 5.4402×10^{10} BTU/CD

ADDITIONAL FUEL GAS MADE = 10,608.6 BPCD (FOE)
= $10{,}608.6 \times 6.35 \times 10^6$
= 6.7365×10^{10} BTU/CD

ADDITIONAL HFO PRODUCTION = 0

ADDITIONAL IMPORTED i-C4 = 1,315.02 BPCD = 21,250x1,315.02 x (349.7x0.565)
= 5.5212×10^9 BTU/CD

ENERGY PENALTY

$$= \frac{5.4402 \times 10^{10} - 6.7365\times10^{10} + 21{,}035.13\times10^6 + 5.5212\times10^9}{3.3849 \times 10^{12}}$$

$= 4.016 \times 10^{-3}$

= 0.40 %

SECTION 4

4.0 COSTS

This section describes the methodology and rationales used to develop the capital and operating cost estimates and associated projections of manpower demands for the engineering, procurement and construction disciplines through to 1985, pertaining to the reducing lead level controls. Appropriate summary tables in addition to relevant graphical representations are included in the following seven subsections.

Based on the computer simulation results of the regional composite refineries reported in Section 3, it has been deduced hypothetically that existing process units in the refineries can cope with the gasoline octane demand induced by imposing either the 2 gm/I.G. or 1 gm/I.G. lead level controls. This does not preclude the possibility that specific refineries may have to inject capital investment to conform with these controls. For this study, it has been assumed that no extra processing capacity is needed. Only the operating costs will change from the Base Case.

However, all regional zero lead level study cases reveal that significant extra processing capacity will be required which will incur considerable expenditure by the refiners.

4.0 COSTS

4.1 CAPITAL COST

The approach adopted to reach the optimal solution during the simulations was one whereby required extra octane barrels were derived principally from the catalytic reformer and the isomerization units, whilst utilizing the existing FCC and associated Polymerization (and Alkylation units) to the best extent possible. Consequently, new precessing capacity and hence capital cost estimation was confined to only Catalytic Reforming, Isomerization Units (and associated HDS units) and Alkylation Units. It should be appreciated that both of these former units operate in a pressurized hydrogen atmosphere and use a platinum catalyst, which can function only with sulphur-free feedstocks. Thus, all naphtha/LSR feedstocks will have to be processed by a Naphtha Desulphurization Unit.

From the computer results, it is manifest that existing HDS processing capacity serving the reformers only is adequate for the zero lead cases except Western region. The reformer capacity is inadequate however due to the applied downrating factor which has been explained earlier. In light of the above, additional HDS capacity has been assumed to be required in association with isomerization unit capacity, and is applicable to the Western zero-lead case only.

For capital cost estimation, the regional new processing capacity as deduced from the computations, was allocated to individual refineries in direct proportion to that refinery's installed reforming capacity expressed as a percentage of the total regional installed reformer capacity. Further, this processing capacity was uprated to reflect conversion from BPCD to BPSD, by application of an operating factor of 90.5% or 330 days/year. Table 4-13 identifies the regional split of processing capacity.

TABLE 4-13

CANADIAN REFINERY CAPACITIES (1979)

Refinery/Location	Crude Distillation	Catalytic & Hydro-cracking		Reforming		Polymerization		Alkylation	
	------- (barrels per calendar day and % of crude capacity) -------								
Golden Eagle-Holyrood	14,000	0	0	1,400	10.0%	0		0	
Gulf Point Tupper	81,000	0	0	10,800	13.3%	0		0	
Imperial-Dartmouth	83,900	24,300	29.0%	8,700	10.4%	2,500	3.0%	0	
Texaco- Halifax	20,000	7,200	36.0%	3,600	18.0%	0		0	
Irving- St. John	250,000	52,000	20.8%	38,500	15.4%	0		0	
ATLANTIC REGION-TOTALS	448,900	83,500	18.6%	63,000	14.0%	2,500	0.6%	0	
BP-Ville D'Anjou	73,000	19,300	26.4%	11,000	15.0%	660	0.9%	0	
Golden Eagle-Quebec	100,000	0	0	15,000	15.0%	0		0	
Gulf-Montreal	77,300	23,500	30.4%	8,600	11.1%	1,000	1.3%	0	
Imperial-Montreal	77,600	22,200	28.6%	10,900	14.0%	2,300	3.0%	0	
Fina-Montreal	95,000	36,145	38.0%	32,000	33.7%	550	0.6%	1,500	1.6%
Shell-Montreal	120,000	36,500	30.4%	23,000	19.2%	0	0	1,600	1.3%
Texaco-Montreal	74,500	26,500	35.6%	12,000	16.1%	1,500	2.0%	3,000	4.0%
QUEBEC-TOTALS	617,400	164,145	26.6%	112,500	18.2%	6,010	1.0%	6,100	1.0%
Texaco-Nanticoke	95,000	40,000	42.1%	24,000	25.3%	0	0	7,400	7.8%
BP-Oakville	80,000	28,100	35.1%	17,850	22.3%	0	0	2,000	2.5%
Gulf-Clarkson	79,100	17,000	21.5%	19,400	24.5%	1,190	1.5%	3,470	4.4%
Imperial-Sarnia	126,100	34,200	27.1%	29,100	23.1%	0	0	5,700	4.5%
Shell-Oakville	44,000	14,000	31.8%	7,000	15.9%	0	0	2,300	5.2%
Shell-Corunna	74,000	23,500	31.8%	23,300	31.5%	1,000	1.35%	0	0
Suncor-Sarnia	90,000	16,100	17.9%	19,750	21.9%	0	0	3,800	4.2%
Petrosar-Sarnia	170,000	0	0	0	0	0	0	0	0
ONTARIO-TOTALS	758,200	172,900	22.8%	140,400	18.5%	2,190	0.3%	24,670	3.2%

TABLE 4-13 (Continued)

CANADIAN REFINERY CAPACITIES (1979)

Consumers-Regina	40,000	13,500	33.7%	9,000	22.5%	1,500	3.75%	0	0
Gulf-Moose Jaw	13,300	0	0	0	0	0	0	0	0
Gulf-Calgary	8,700	0	0	0	0	0	0	0	0
Gulf-Edmonton	80,000	28,800	36.0%	13,000	16.2%	0	0	9,000	11.2%
Husky-Lloydminster	12,000	0	0	0	0	0	0	0	0
Imperial-Strathcona	150,700	43,700	29.0%	21,300	14.1%	0	0	9,700	6.4%
Shell-St. Boniface	30,000	9,000	30.0%	8,200	27.3%	500	1.7%	0	0
Shell-Bowden	5,000	0	0	2,500	50.0%	0	0	0	0
Texaco-Edmonton	24,000	9,000	37.5%	4,500	18.7%	900	3.7%	0	0
Petrocan-Taylor	14,300	5,000	35.0%	2,050	14.3%	0	0	930	6.5%
Chevron-Burnaby	35,000	8,000	22.9%	11,700	33.4%	600	1.7%	0	0
Gulf-Port Moody	37,200	11,400	30.6%	9,700	26.1%	0	0	1,440	3.9%
Gulf-Kamloops	9,300	2,790	30.0%	1,620	17.4%	220	2.4%	0	0
Husky-Prince George	10,000	3,000	30.0%	1,400	14.0%	0	0	0	0
Imperial-Ioco	40,700	12,300	30.2%	6,200	15.2%	1,000	2.4%	0	0
Imperial-Norman Wells, N.W.T.	2,900	0	0	0	0	0	0	0	0
Shell-Burnaby	22,000	6,000	27.3%	2,700	12.3%	500	2.3%	0	0
WESTERN REGION-TOTALS (1)	535,100	152,490	28.5%	93,870	17.5%	5,220	1.0%	21,070	3.9%

Note (1) Does not include locally consumed light distillates from Syncrude and Suncor in Ft. McMurray, Alberta.

Source : Petroleum Refineries in Canada, December 1979, issued by Energy, Mines and Resources, Canada Energy Policy Sector.

TABLE 4-1

SUMMARY OF REFORMER CAPACITIES FOR PRODUCING LEAD-FREE GASOLINES

REGION	Installed 1979 98 RON BPCD	Equivalent Installed 103 RON BPCD	Required From Simulations BPCD	Incremental BPCD	For Cost Estimation BPSD
ATLANTIC	63,000	0	38,000	38,000	42,677
QUEBEC	112,500	0	69,600	69,600	76,982
ONTARIO	140,400	0	75,500	75,500	83,508
WESTERN	93,920	70,440 (@ 100 RON)	94,000	23,560	26,059

SUMMARY OF ISOMERIZATION/HF ALKYLATION CAPACITIES REQUIRED FOR THE PRODUCTION OF LEAD FREE GASOLINES

REGION	ISOMERIZATION CAPACITY		HF ALKYLATION CAPACITY	
	BPCD	For Cost Estimation BPCD	BPCD	For Cost Estimation BPSD
ATLANTIC	12,837	14,417	1,746	1,961
QUEBEC	21,853	24,171	1,387	1,534
ONTARIO	38,320	42,384	4,216	4,663
WESTERN	36,938	40,856 *	---	---

* Extra HDS capacity is required for desulphurization of LSR for Western region only i.e. 40856 BPCD.

Table 4-1 shows the regional processing requirements for the zero lead case.

The associated capital costs were developed on the basis of the following assumptions:

Catalytic Reforming	- bi-metallic (platinum/rhenium) catalyst, semi-regenerative and continuous type, desulphurized naphtha feedstock.
Isomerization	- LSR C_5/C_6 feedstock, platinum catalyst, no recycle
HDS	- LSR C_5/C_6 feedstock
Alkylation	- HF Alkylation

Capital cost data was derived primarily from major licensors of the proprietary technology during the course of the study to ensure that latest data was applied. This information was supplemented as required with 'in-house' refinery planning and experience data obtained from our cost library.

Pertinent cost estimating data was complied and a representative average cost curve developed for varying processing capacities of the three basic types of units considered. These curves are appended to this section. (See Figures 4-1 to 4-3 and 4-16 to and 4-17.)

The capital costs so developed can be classified as "Order of Magnitude" estimates and are normally considered to have a ± 30% accuracy pertaining to a nominal unit capacity. The scope of the estimate includes all fixed capital costs which encompass:-

- Battery Limit Process Units
- Site Preparation
- Industrial Buildings
- Storage Facilities
- Auxiliary Facilities
- Offsite Piping and Services

Land development costs have been excluded.

The data from the cost library is classified and grouped into complete cost units which include all direct and indirect costs, necessary to engineer, procure and construct all battery limit and offsite equipment and material. The grouping is with respect to:-

- Equipment
- Material and Labour
- Indirects

The Equipment component includes the cost of major process items such as towers, reactors, exchangers and offsite storage, utility plants and auxiliaries. The Material dollar includes all necessary items for piping, instrumentation, civil structural and electrical systems. Labour is comprised of direct field costs and construction overheads related to aforementioned equipment and material.

The indirect cost component covers services provided by engineering, procurement, field supervision, fees and the project management team.

Estimate Structure

The total capital investment cost is given per refinery location. The estimate structure of each process unit comprises:

New Equipment ____________
Material ____________
Offsites ____________
Direct Cost __________________________ ______
Engineering __________________
Total Capital Cost ____________________________
Catalyst ____________
Total Investment Cost ============

One or more of these process units is the total investment cost of a refinery.

Pricing Basis

The cost-capacity curves so derived are represented by:-

$$\text{Canital Cost} = \left(\frac{\text{Capacity in Study Case}}{\text{Capacity in Reference}}\right)^{Y} \times \text{Reference Cost}$$

where the exponent 'Y' is the scale factor. This method has worked well historically where capacities vary no more than three times.

However, in several instances, the increased requirements resulted in only marginal capacity increases. Consequently the ratio of "capacity study" vs "capacity reference" increased to more than three (3). This "bottom limit" appears to be a correct reflection of the relative feasibility of such small plants; available historical data consists of only much larger plants.

To fully consider these low capacities, it would be necessary to address new technologies, the possibilities of revamps at existing plants or even their reconstruction to higher capacities. This, however, may not be feasible, or as in the case of a revamp, even possible. At any rate, such discussion is beyond the scope of this study.

Consequently, it was necessary to make some adjustment to the methodology as it applies to these "bottom limit" cases. For each process, a range of capacities was established based on the characteristic cost vs. capacity relationship. All plants whose capacities fell below the lower boundary of this range were costed at this point, while those that occurred within the range had contingency factor applied to them. This adjustment is based on the assumption, that as capacities decrease, the corresponding investment densities increases.

In graphic terms this means, that the "cost vs. capacity" curve levels off below the "bottom limit". Therefore, the relationship between investment costs and unit capacity otherwise clearly evident in the cases of plants with higher capacities, becomes somewhat - though not incorrectly - distorted.

Relative fractional cost of total offsite investment was prepared and used in this report for power plants, utility plants, auxiliary equipment, site improvements, distribution and supply systems. On major offsite costs, such as tank farms, export terminals, main headers and blending systems, a general scope was prepared and costed. These costs were adjusted to facilitate the unique requirements of each refinery location.

Appropriate construction parameters pertaining to each refinery location across Canada were applied to the cost estimate. Adjustments are generally labour related and include geographical, historical, economic and climate conditions. These adjustments have been made in the form of relative factors, backed by known, on-hand cost comparisons and published statistics.

Other Qualifications

Validity date of all costs in this refinery study is fourth (4th) quarter of 1979. The accuracy of estimates of total capital investments is in the 'order of magnitude' range.

Inclusions in Estimates:

- Permits, Authorizations, Royalties, etc.
- Taxes, Duty
- Spare Parts
- Catalysts
- On and Off-Site Facilities

Refinery overhead costs and land development costs were excluded, in addition to interest charges and costs incurred by delays, strikes, premium time and forecasted manpower shortage.

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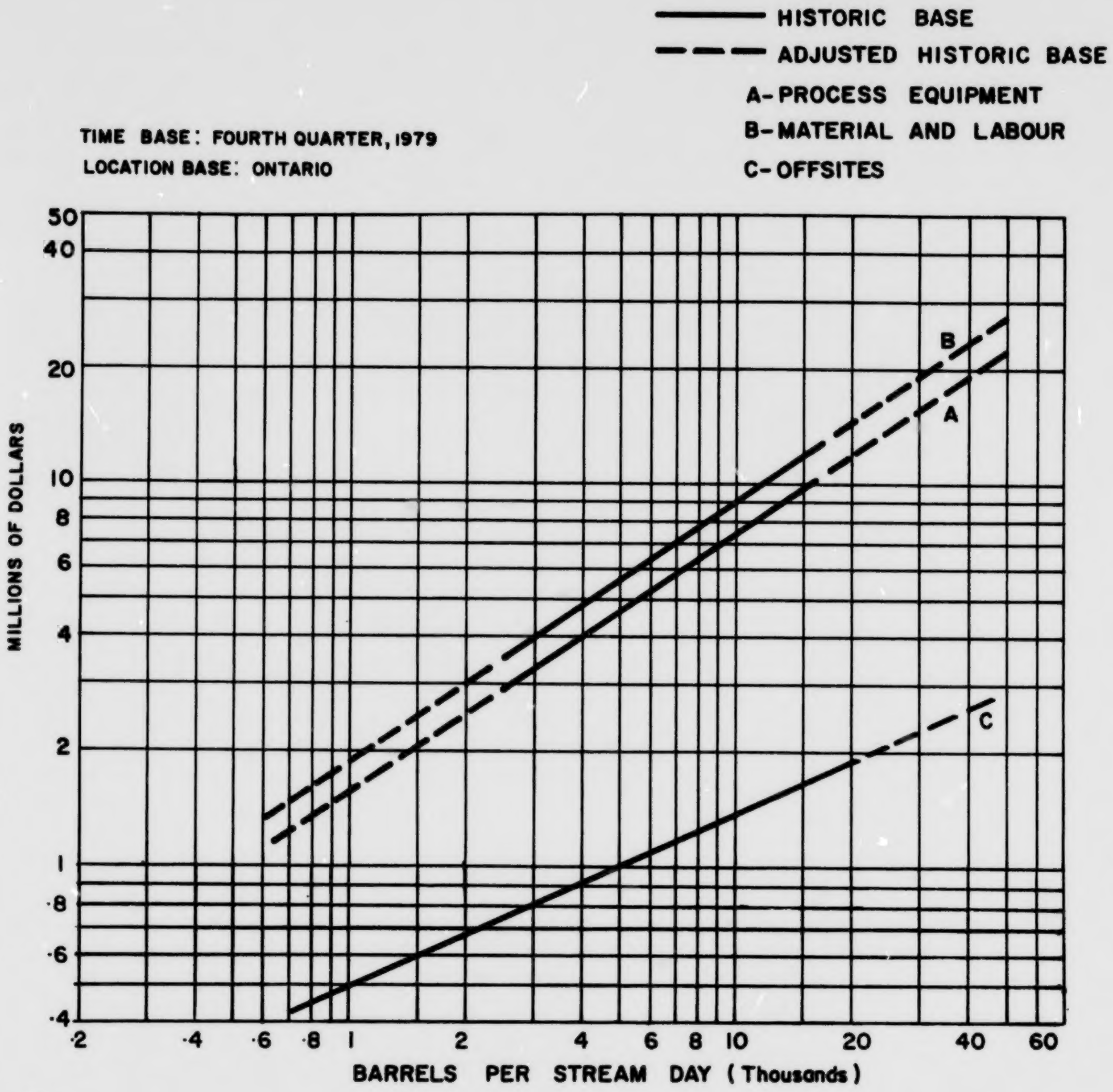
Capital Cost Summaries

The Summary Tables appended to this section report the estimated capital investment costs by region. Additionally, a complete breakdown of costs pertaining to each refinery is provided, although the refinery locations are coded purposely for obvious reasons.

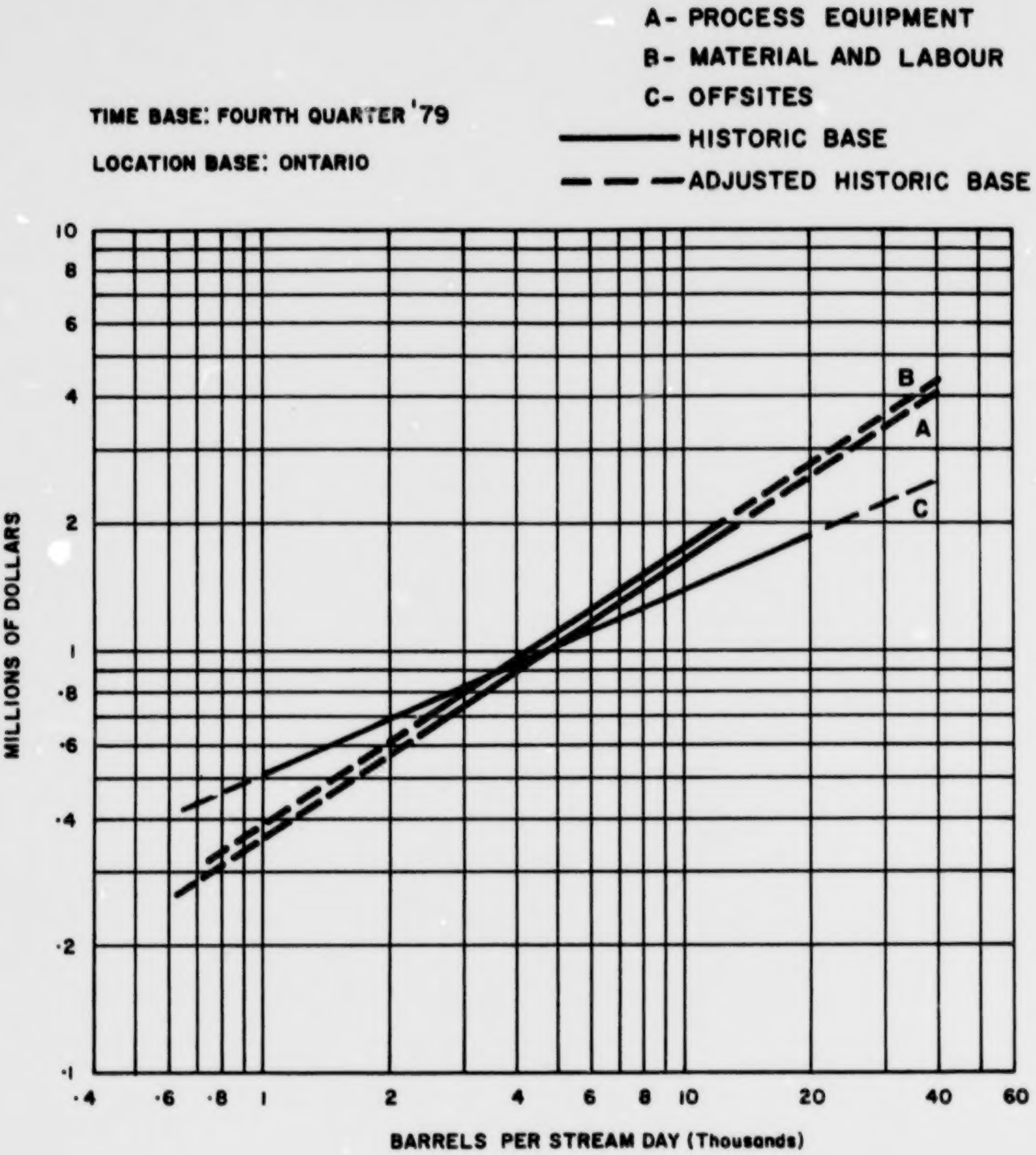
Cost figures are reported in 1979 Canadian Dollars.



DIRECT COST FOR REFORMING

Semi Regenerative

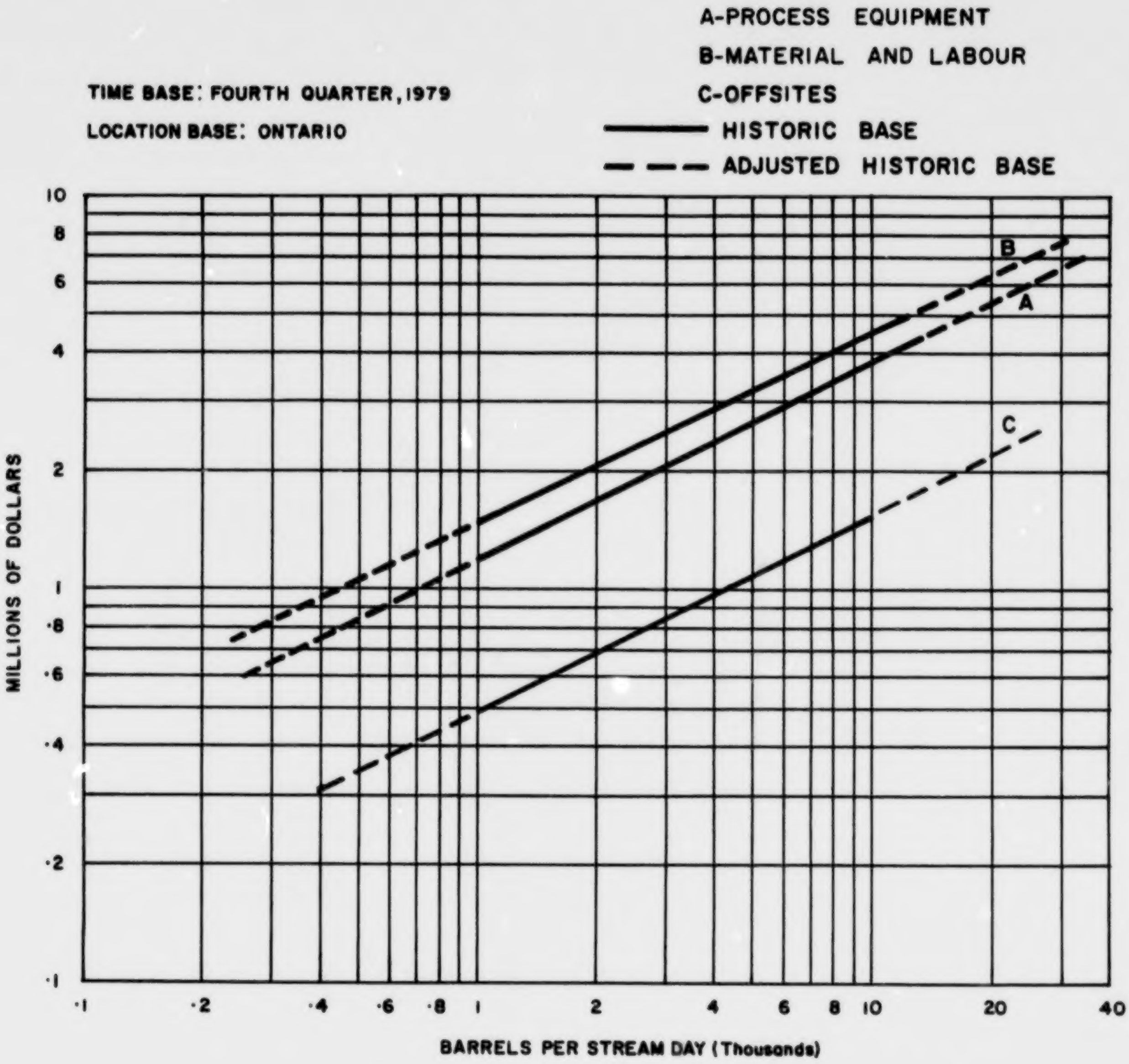
FIGURE 4·1



DIRECT COST FOR ISOMERIZATION

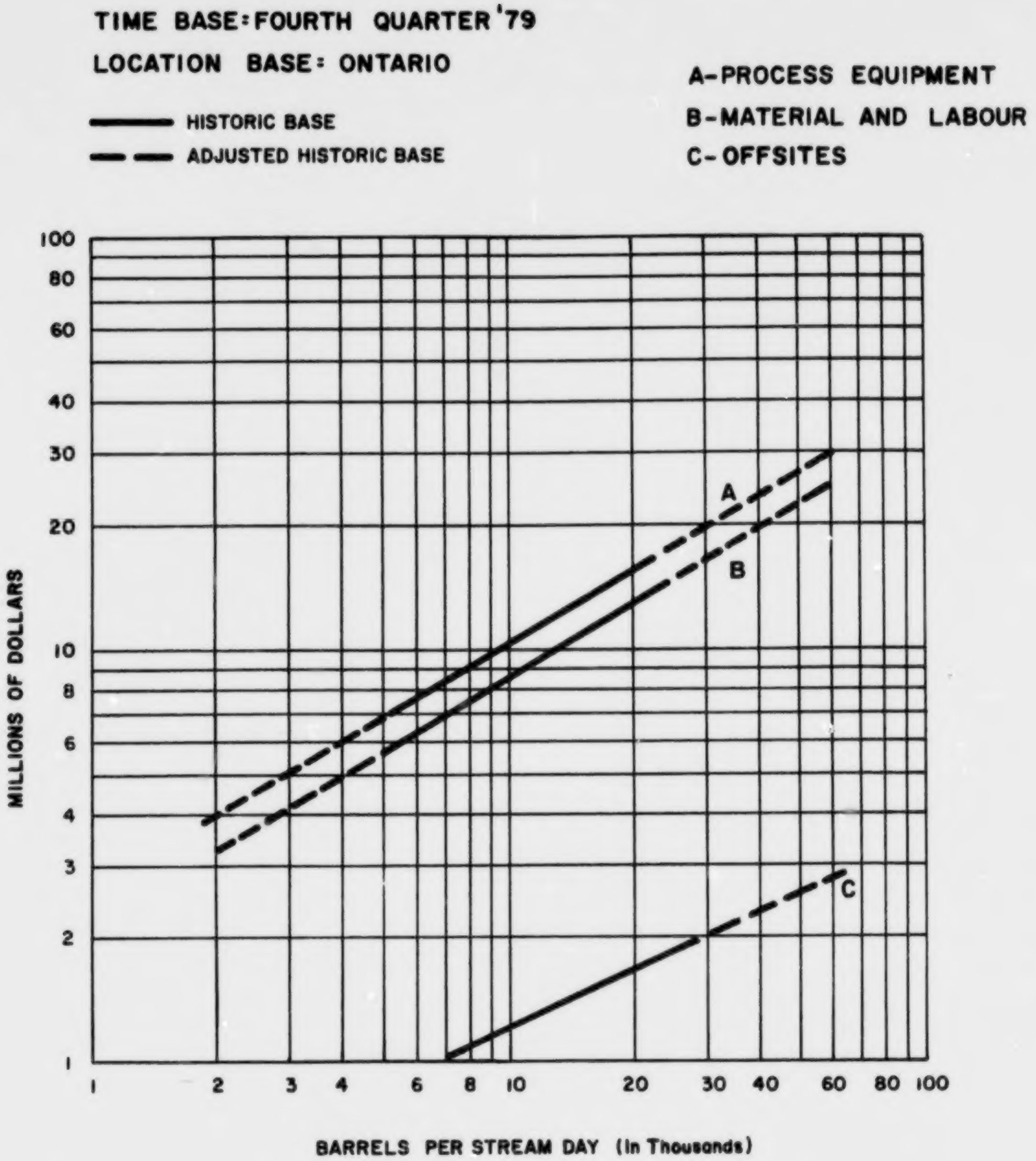
LSR Straight Run

FIGURE 4·2



DIRECT COST FOR HYDROTREATING NAPHTHA

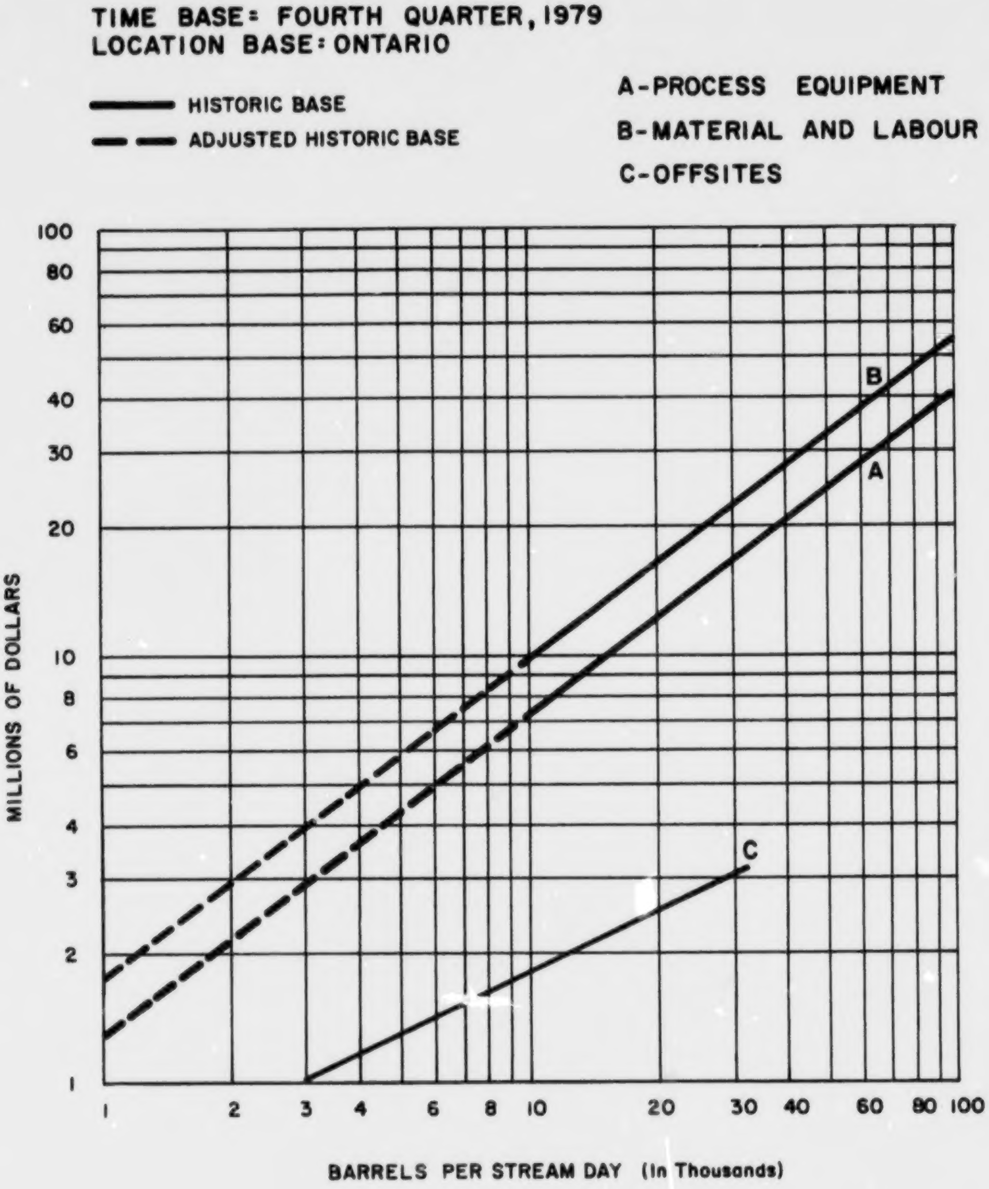
FIGURE 4·3



DIRECT COST FOR

CONTINUOUS CATALYTIC REFORMING PLANT

FIGURE 4·16



DIRECT COST FOR

HF ALKYLATION PLANT

FIGURE 4·17

4.0 COSTS

4.2 OPERATING COSTS

Variations in refinery operating costs for each reducing lead level control have been calculated for each region. Basically, these costs have been derived from an analysis of nine basic categories, namely:-

1. Power
2. Fuel and Steam
3. Additional crude oil
4. Imported I-C_4
5. Credit for extra HFO production
6. Catalysts Replacement
7. Wages for extra Refinery Personnel
8. Credit for Reduced Lead Additive Consumption
9. Non-lead Additive Consumption

Table 2-17 lists the pertinent data developed for regional feedstock, product and utility costs.

Fuel and Steam Costs

The cost of steam was taken to be 1.5 times the HFO cost for the steam energy equivalent. All fuel used for processing requirements is assumed to be supplied primarily by the fuel gas generated within the refinery and supplemented by heavy fuel oil if necessary. Any excess of fuel gas will probably be condensed and sold as LPG by the refineries. However, for calculation and evaluation purposes, the price of an equivalent quantity of HFO is credited.

Catalyst Replacement Costs

The catalyst replacement cost calculations are based on the capacities of the additional reformer, isomerization and HDS units. A bi-metallic catalyst containing platinum and rhenium supported on silica base is used for the reformer. The cost is taken to be one-third of the initial catalyst charge spread over one-and-one-half years. Catalyst requirements are much less for the continuous type reformer versus the semi-regenerative type.

Extra Refinery Personnel

For each refinery, the following additional personnel are considered necessary to operate the additional reformer and isomerization/HDS process units required in the production of a total lead-free gasoline pool, irrespective of their capacities:-

	No./Shift	No. Per Day
Process Operators	4 (Note 1)	18 (Note 2)
Laboratory Technician		1
Utility Operator		1
Maintenance Engineer		1
Clerk		1
Process Engineer		1
		23 (Note 3)

Note 1: 5 if alkylation unit included
Note 2: 22 " " " "
Note 3: 27 " " " "

The process operators are responsible for the day-to-day operation and control of the process units, and they will be stationed in the process control room. An extra laboratory technician is deemed necessary to conduct routine sample analysis to ensure product quality control. The additional utility operator will be responsible for proper and adequate supply of utilities such as - water, steam, and chemicals, etc.. A maintenance engineer has been allowed for, to perform minor repairs to mechanical/electrical/instrumentation equipment on a daily basis.

The extra clerk would supplement the office staff to perform routine administrative and clerical duties. The process engineer would oversee and co-ordinate the new process units on a daily basis. He would monitor process plant operation, conduct test runs, and write technical reports to management suggesting optimum operational modes.

The qualifications needed for these personnel are summarized below:-

Personnel Category	Qualifications
Process Operators	- High School Education + practical engineering apprenticeship
Utility Operator	- High School Education + practical engineering apprenticeship
Maintenance Engineer	- High School Education + practical engineering apprenticeship
Laboratory Technician	- High School Education + chemistry bent
Clerk	- High School Education
Process Engineer	- Degree in Chemical Engineering

The process operators would be on shift work whereas all other personnel on the list would operate on a daily basis. The average remuneration for each personnel classification has not been subdivided due to lack of readily available data. Statscan published an average figure relating to process type operators and it is considered fairly representative of the average remuneration applicable to the general classifications mentioned above. This figure is $95.632/day per person.

Note that it takes the equivalent of 4½ men on the payroll for each shift job to cover vacations, statutory holidays and illness, etc..

The control and operation of the new process units is assumed to be integrated and they would be physically located in one area of the refinery. The personnel requirements are considered to remain the same with the addition of isomerization, HDS, and/or reformer units, for the purposes of the study. However, a new alkylation unit will require an operator, additional to the figures mentioned above, raising the total additional refinery personnel to 27, for each refinery.

The estimated total requirement across Canada for extra refinery personnel is between 900 to 1000, and the regional distribution is given below:-

Category	Atlantic	Quebec	Ontario	Western
Process Operators	115	161	186	306
Laboratory Tech.	5	7	8	17
Utility Operators	5	7	8	17
Maintenance Eng.	5	7	8	17
Clerk	5	7	8	17
Process Engineer	5	7	8	17
	140	196	224	391

The total process incremental operating cost variations were calculated for each case and all results are reported in Table 4-2 to 4-5. These figures exclude any capital depreciation. Table 4-11 summarizes the overall incremental operating costs to produce lead-free gasoline expressed in cents/litre.

TABLE 4-2

SUMMARY OF OPERATING COSTS BY REGION

ATLANTIC

	2 LEAD $'000/CD	1 LEAD $'000/CD	NO LEAD $'000/CD
1. POWER COST	0.417	1.235	2.476
2. FUEL & STEAM COSTS	-15.728	-24.383	-41.651
3. COST OF ADDITIONAL CRUDE	34.098	63.791	76.146
4. COST OF IMPORTED i-C4	-	-	-
5. CREDIT FOR EXTRA HFO PROD'N	-	-	-
6. CATALYSTS REPLACEMENTS COSTS	-	-	2.156
7. WAGES	-	-	12.142
8. CREDIT FOR LEAD ADDITIVES	-2.491	-6.019	-9.547
9. COST OF MMT			1.742
TOTAL	16.296	34.624	43.464

TABLE 4-3

SUMMARY OF OPERATING COSTS BY REGION

QUEBEC REGION

	2 LEAD $'000/CD	1 LEAD $'000/CD	NO LEAD $'000/CD
1. POWER COST	-0.125	- 2.282	1.236
2. FUEL & STEAM COSTS	-36.774	-77.843	-87.253
3. COST OF ADDITIONAL CRUDE	64.815	184.779	144.205
4. COST OF IMPORTED i-C4	-	-	-
5. CREDIT FOR EXTRA HFO PROD'N	-	-	-
6. CATALYSTS REPLACEMENTS COSTS	-	-	2.207
7. WAGES	-	-	16.541
8. CREDIT FOR LEAD ADDITIVES	-4.722	-10.775	-16.828
9. COST OF MMT	-	-	6.684
TOTAL	23.194	93.879	66.792

TABLE 4-4

SUMMARY OF OPERATING COSTS BY REGION

ONTARIO

	2 LEAD $'000/CD	1 LEAD $'000/CD	NO LEAD $'000/CD
1. POWER COST	-	-1.928	1.169
2. FUEL & STEAM COSTS	-	-77.471	-135.594
3. COST OF ADDITIONAL CRUDE	-	112.788	141.526
4. COST OF IMPORTED i-C4	-	-19.104	52.549
5. CREDIT FOR EXTRA HFO PROD'N	-	-	-
6. CATALYSTS REPLACEMENTS COSTS	-	-	2.883
7. WAGES	-	-	19.122
8. CREDIT FOR LEAD ADDITIVES	-	-6.959	-16.532
9. COST OF MMT	-	-	3.403
TOTAL	-	7.326	68.526

TABLE 4-5

SUMMARY OF OPERATING COSTS BY REGION

WESTERN

	2 LEAD $'000/CD	1 LEAD $'000/CD	NO LEAD $'000/CD
1. POWER COST	-	1.603	3.898
2. FUEL & STEAM COSTS	-	- 11.630	- 93.214
3. COST OF ADDITIONAL CRUDE	-	44.316	161.439
4. COST OF IMPORTED i-C4	-	- 22.768	- 48.688
5. CREDIT FOR EXTRA HFO PROD'N	-	-	-
6. CATALYSTS REPLACEMENTS COSTS	-	-	2.408
7. WAGES	-	-	37.392
8. CREDIT FOR LEAD ADDITIVES	-	- 11.175	- 22.430
TOTAL	-	0.346	40.805

Total Incremental Operating Costs

The incremental operating cost incurred by the reduction of lead levels in gasolines consists of the following:-

1. Capital Charges
2. Insurance
3. Local Taxes
4. Maintenance
5. Process Operating Cost

The first four items are a function of the plant investment and are usually expressed as a fixed percentage of the total capital cost. The total capital cost is the sum of the capital equipment cost and the crude inventory charge (based on a 30 day inventory requirement of the incremental crude capacity). The plant investment is assumed to be recovered over four years on a straight-line basis. (i.e. annual capital charges allowance equals 25% of the total capital cost and represents the assumed before income tax return to the refiners, including provision for depreciation). Insurance on equipment usually approximates to 0.5% of the capital cost per year. Local taxes are assumed to be 1% of the total capital cost per year. Maintenance includes material and labor, and has been estimated to be 5.5% of total capital cost.

TABLE 4-6
ANNUAL OPERATING COST
ATLANTIC

FOR LEAD-FREE GASOLINE PRODUCTION

CAPITAL EQUIPMENT COST	=	$ 177,055,000
CRUDE INVENTORY	=	$ 2,284,380
TOTAL CAPITAL COST	=	$ 179,339,380

ANNUAL OPERATING COST $/YR

		2 LEAD	1 LEAD	0 LEAD
CAPITAL CHARGES	(25%)	-	-	44,834,845
INSURANCE	(0.5%)	-	-	896,697
LOCAL TAXES	(1.0%)	-	-	1,793,394
MAINTENANCE	(5.5%)	-	-	9,863,666
PROCESS OPERATING COST		5,948,040	12,637,760	15,864,360
TOTAL		5,948,040	12,637,760	73,252,962

TABLE 4-7
ANNUAL OPERATING COST
QUEBEC

FOR LEAD-FREE GASOLINE PRODUCTION

CAPITAL EQUIPMENT COST = 200,105,000
CRUDE INVENTORY = 4,326,150

TOTAL CAPITAL COST = 204,431,150

ANNUAL OPERATING COST

		$/YR 2 LEAD	1 LEAD	0 LEAD
CAPITAL CHARGES	(25%)	-	-	51,107,788
INSURANCE	(0.5%)	-	-	1,022,156
LOCAL TAXES	(1.0%)	-	-	2,044,312
MAINTENANCE	(5.5%)	-	-	11,243,713
PROCESS OPERATING COST		8,465,810	34,265,835	24,379,080
TOTAL		8,465,810	34,265,835	89,797,049

TABLE 4-8
ANNUAL OPERATING COST
ONTARIO

FOR LEAD-FREE GASOLINE PRODUCTION

CAPITAL EQUIPMENT COST = $ 255,847,000
CRUDE INVENTORY = $ 4,245,780

TOTAL CAPITAL COST = $ 260,092,780

ANNUAL OPERATING COST

		$/YR 1 LEAD	0 LEAD
CAPITAL CHARGES	(25%)	-	65,023,195
INSURANCE	(0.5%)	-	1,300,464
LOCAL TAXES	(1.0%)	-	2,600,928
MAINTENANCE	(5.5%)	-	14,305,103
PROCESS OPERATING COST		2,673,990	25,011,990
TOTAL		2,673,990	108,241,680

TABLE 4-9
ANNUAL OPERATING COST
WESTERN

FOR LEAD-FREE GASOLINE PRODUCTION

CAPITAL EQUIPMENT COST	=	$253,210,000
CRUDE INVENTORY	=	4,843,163
TOTAL CAPITAL COST	=	$258,053,163

ANNUAL OPERATING COST

		$/YR 1 LEAD	0 LEAD
CAPITAL CHARGES	(25%)	-	64,513,291
INSURANCE	(0.5%)	-	1,290,266
LOCAL TAXES	(1.0%)	-	2,580,532
MAINTENANCE	(5.5%)	-	14,192,924
PROCESS OPERATING COST		126,290	14,893,825
TOTAL		126,290	97,470,838

TABLE 4 -10

ANNUAL OPERATING COST

CANADA

FOR LEAD-FREE GASOLINE PRODUCTION

Capital Equipment Cost	=	$886,217,000
Crude Inventory	=	$ 15,699,473
Total Capital Cost	=	$901,916,473

ANNUAL OPERATING COST $/YR

		2 LEAD	1 LEAD	0 LEAD
Capital Charges	(25%)	-	-	225,479,119
Insurance	(0.5%)	-	-	4,509,583
Local Taxes	(1.0%)	-	-	9,019,166
Maintenance	(5.5%)	-	-	49,605,406
Process Operating Cost		14,413,850	49,703,875	80,149,255
Total		14,413,850	49,703,875	368,762,529

TABLE 4-11

SUMMARY OF REGIONAL INCREMENTAL GASOLINE COSTS

MOTOR GASOLINE PRODUCTION

	ATLANTIC	QUEBEC	ONTARIO	WESTERN	CANADA
L.R. (BPCD)	50073	85908	135858	159718	431557
U.R. (BPCD)	20630	40288	62423	58307	181648
U.P. (BPCD)	6619	32044	18213	13350	70226
	77322	158240	216494	231375	683431

INCREMENTAL ANNUAL OPERATING COST

	ATLANTIC	QUEBEC	ONTARIO	WESTERN	CANADA
2 LEAD $/YR	5,948,040	8,465,810	-	-	14,413,850
1 LEAD $/YR	12,637,760	34,265,835	2,673,990	126,290	49,703,875
0 LEAD $/YR	73,252,962	89,797,049	108,241,680	97,470,838	368,762,529

INCREMENTAL COST IN ¢/1 (BASED ON ALL MOTOR GAS)

	ATLANTIC	QUEBEC	ONTARIO	WESTERN	CANADA
2 LEAD	0.13	0.09	-	-	0.11
1 LEAD	0.28	0.37	0.02	0	0.13
0 LEAD	1.63	0.98	0.86	0.73	0.93

INCREMENTAL COST IN ¢/1 (BASED ON LEADED REG. ONLY)

	ATLANTIC	QUEBEC	ONTARIO	WESTERN	CANADA
2 LEAD	0.20	0.17	-	-	0.18
1 LEAD	0.43	0.69	0.03	0	0.20
0 LEAD	2.52	1.80	1.37	1.05	1.47

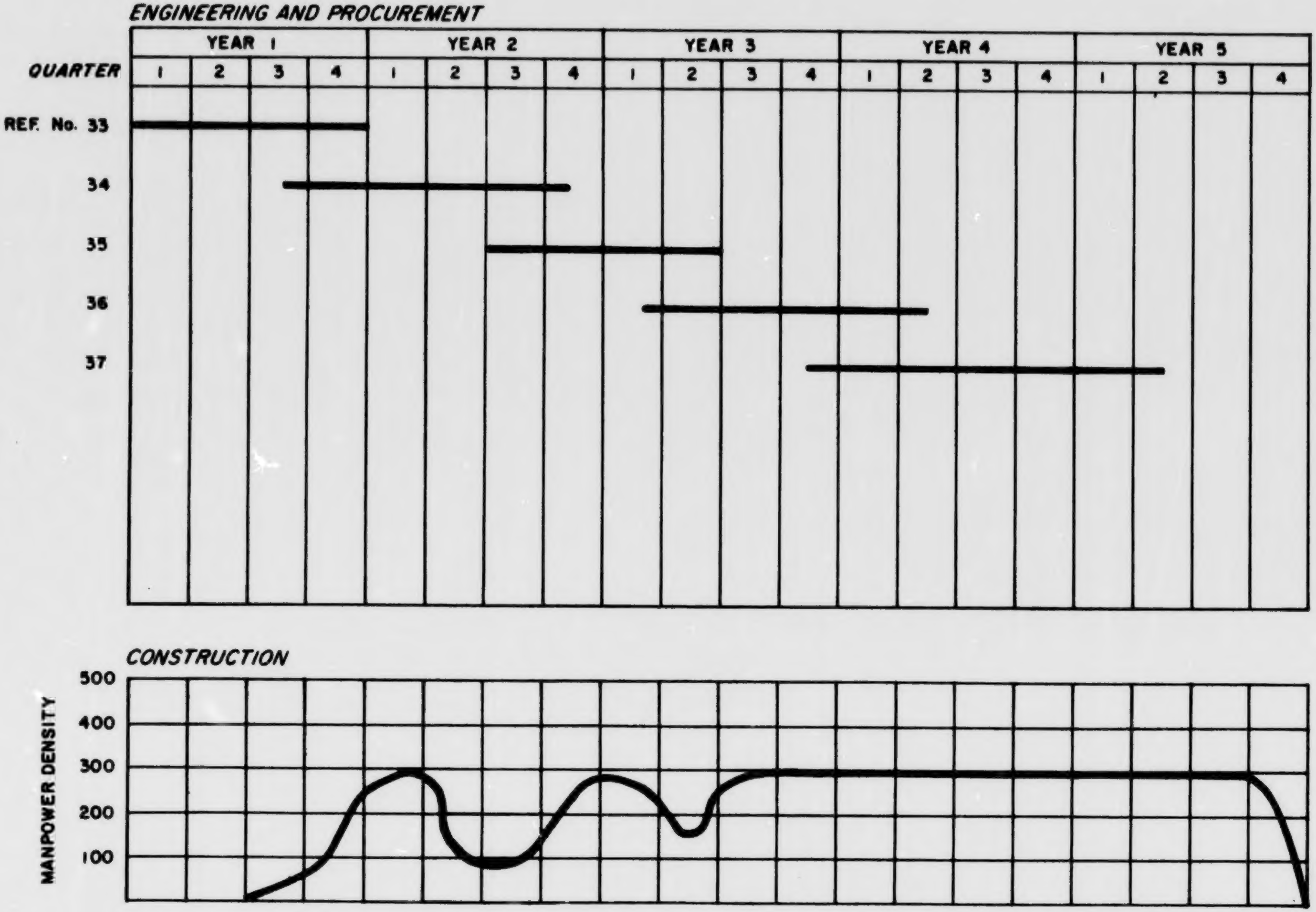
4.0 COSTS

4.3 PROJECT CONSTRUCTION SCHEDULES

These Schedules have been developed from the total "Forecasted Labour Manhours" for the field construction phase of the Lead Reduction Programme estimate as defined in Section 4.1. The total field labour manhours, assumed to remain constant, will establish the approximate manpower density and field duration. These two variables are determined from normalized typical distribution profiles ("S" curve and histogram), based on actual construction cost records and plotted on a man/month scale. The sum of the area under the resultant manning profile is equal to the "Total Field Labour Manhours" in the capital investment estimate of this study.

The concept of this Schedule is concerned only with construction cash flows. The objective will be to recover outlays and expenditures as soon as possible. Any deferments due to economic or other conditions have not been considered.

The Schedules reflect an assumed engineering/procurement lead time of 6-12 months and a construction phase lasting approximately 10-15 months. The refiners themselves will need an additional minimum of six months approximately to develop optimum processing schemes for the individual refineries.

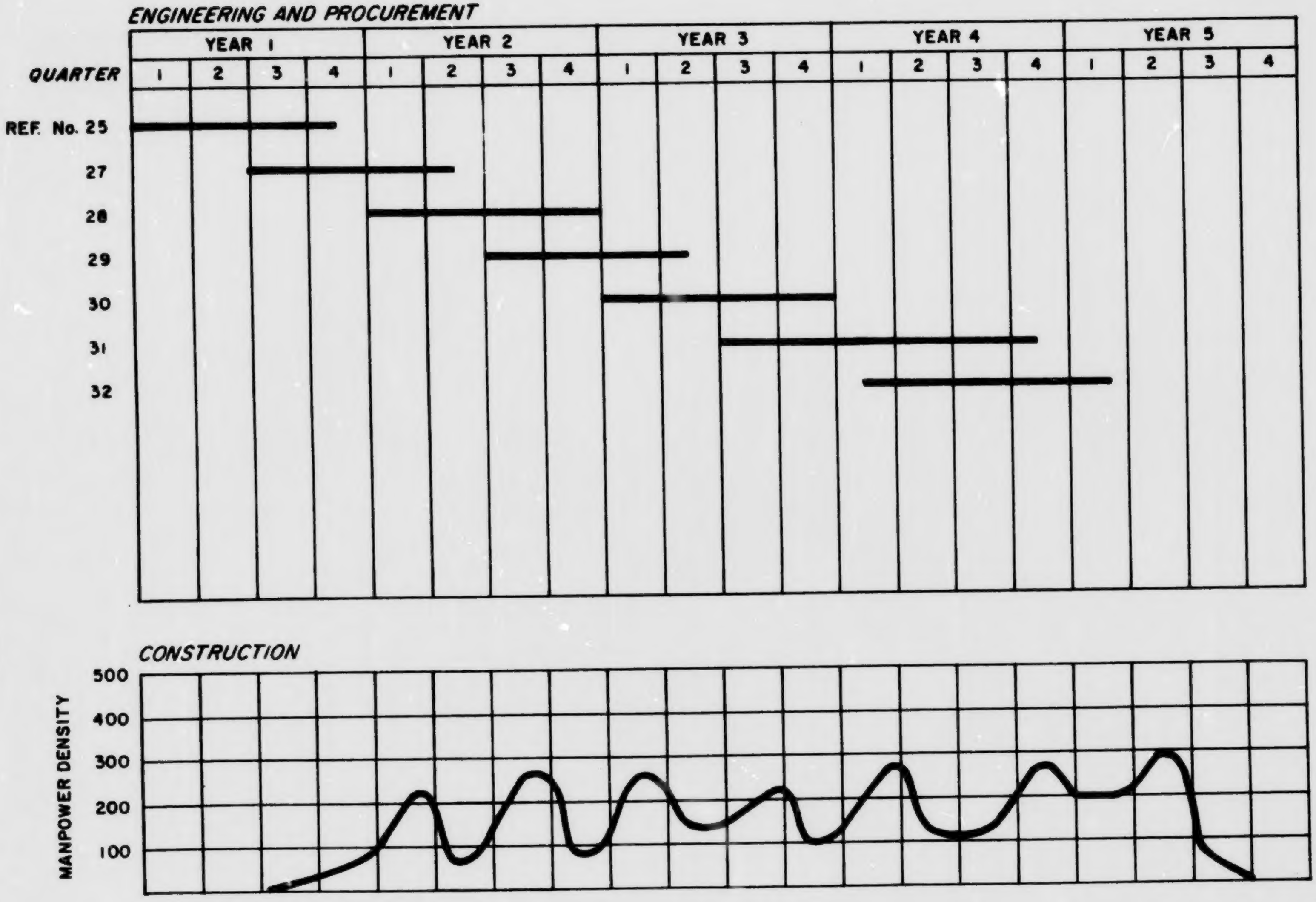


ATLANTIC REGION

ENGINEERING, PROCUREMENT AND CONTRUCTION

MANPOWER SCHEDULED PROJECTIONS

FIGURE 4·4

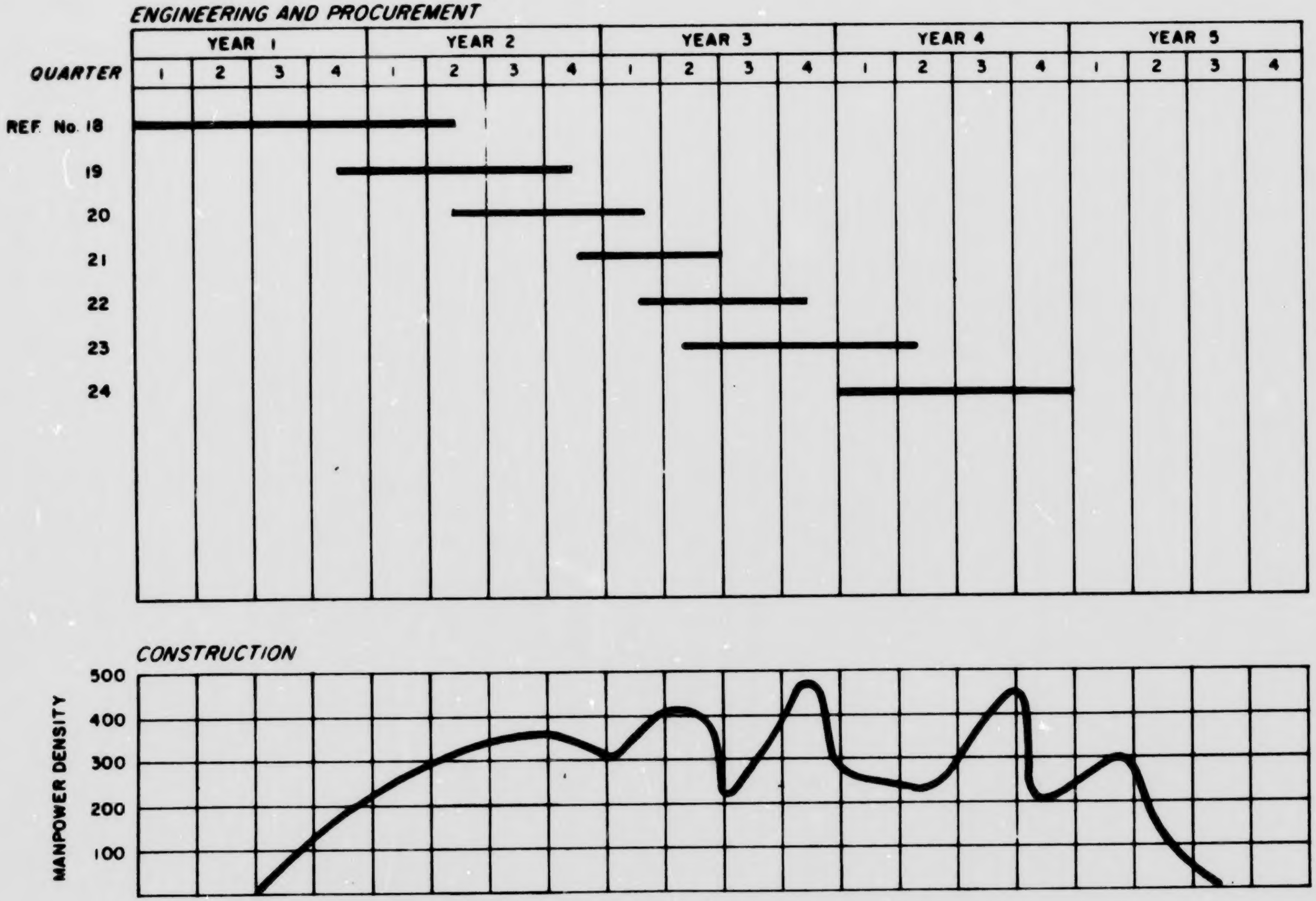


QUEBEC REGION

ENGINEERING, PROCUREMENT AND CONTRUCTION

MANPOWER SCHEDULED PROJECTIONS

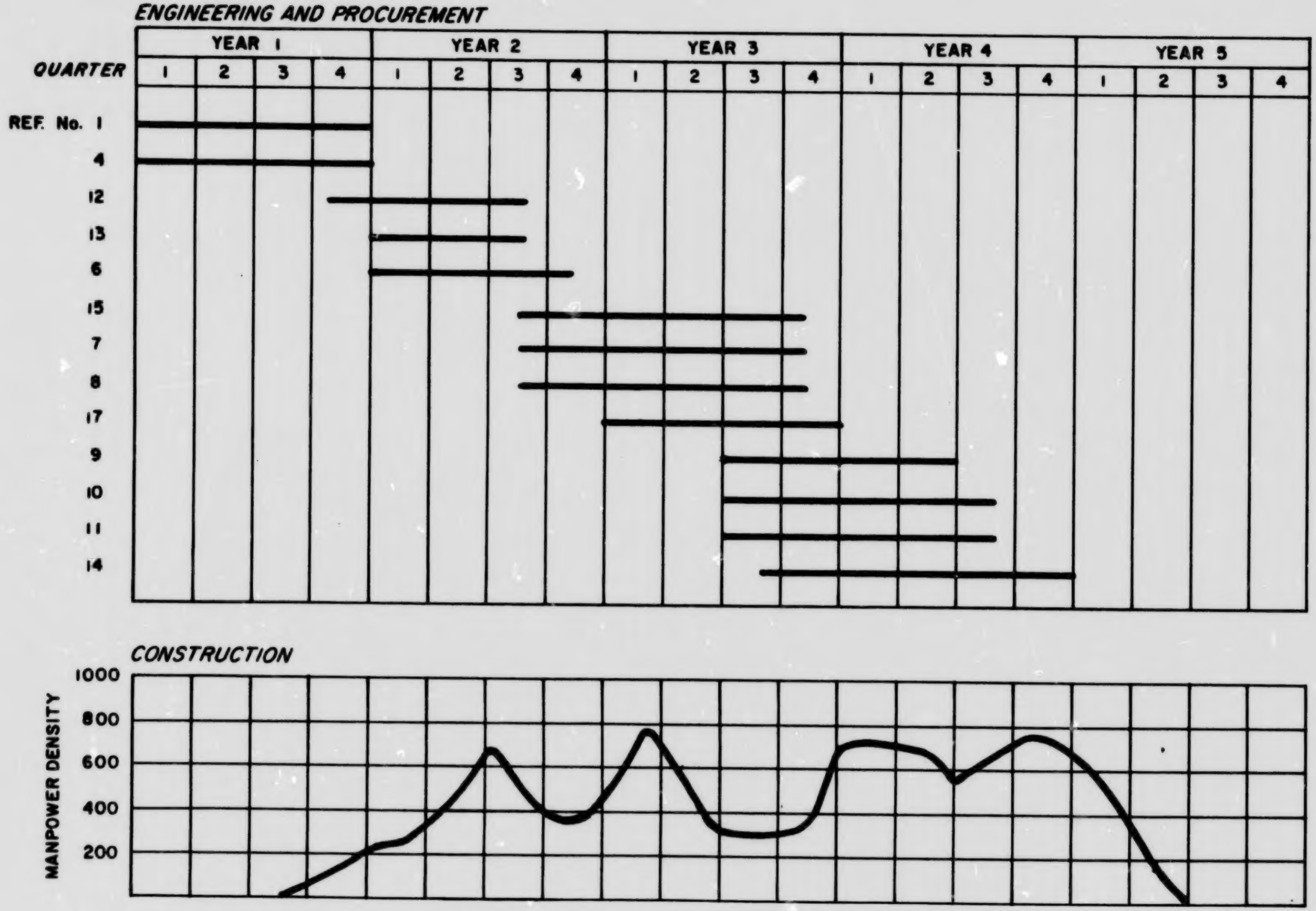
FIGURE 4·5



ONTARIO REGION

ENGINEERING, PROCUREMENT AND CONTRUCTION MANPOWER SCHEDULED PROJECTIONS

FIGURE 4·6



WESTERN REGION

ENGINEERING, PROCUREMENT AND CONTRUCTION

MANPOWER SCHEDULED PROJECTIONS

FIGURE 4·7

4.0 COSTS

4.4 EQUIPMENT DELIVERY INFORMATION

Delivery times for equipment in 1980 averaged 6 months, but are not expected to remain at this level. 12 to 18 months periods can be expected in 1981 for centrifugal compressors, and steam turbine drivers. After 1981, orders created by major development projects can be expected to extend deliveries considerably.

The "1980 Canadian Cost Survey"[1] by ESSO Resources Canada Limited reports the following delivery information:

FABRICATION SHOP LOADINGS FOR 1980

	% OF SHOP CAPACITY BEING UTILIZED
Pressure Vessels	70
Heat Exchangers	65
Furnaces	61
Pipe Fabrication	60
Structural Steel	61
Pumps	74
Compressors	64
Instruments	69
Electrical	83

---

1. 1980 Canadian Process Plant Cost Survey, ESSO Resources Canada Limited.

As the projected rate of growth of energy related industries increases, the fabrication shop loading can reasonably be expected to reach its full present potential, other economic sectors notwithstanding. However, shortages of equipment and materials are not anticipated; Canada's manufacturing sector appears to have the capacity to accommodate the projected growth. Nevertheless, longer delivery periods, averaging 8 to 12 months, can be expected, with up to 2 year periods for aforementioned specific items.

4.0 COSTS

4.5.1 CONSTRUCTION MANPOWER FORECASTS

Labour supply will be an increasing problem in the forecasted period of this report. Using the known, projected "Capital Investments" for planned refining, petrochemical and related industries (from published periodicals)[1], the construction manpower density can be forecasted and summarized in the following table No. 4-12 and figures 4-8 to 4-15.

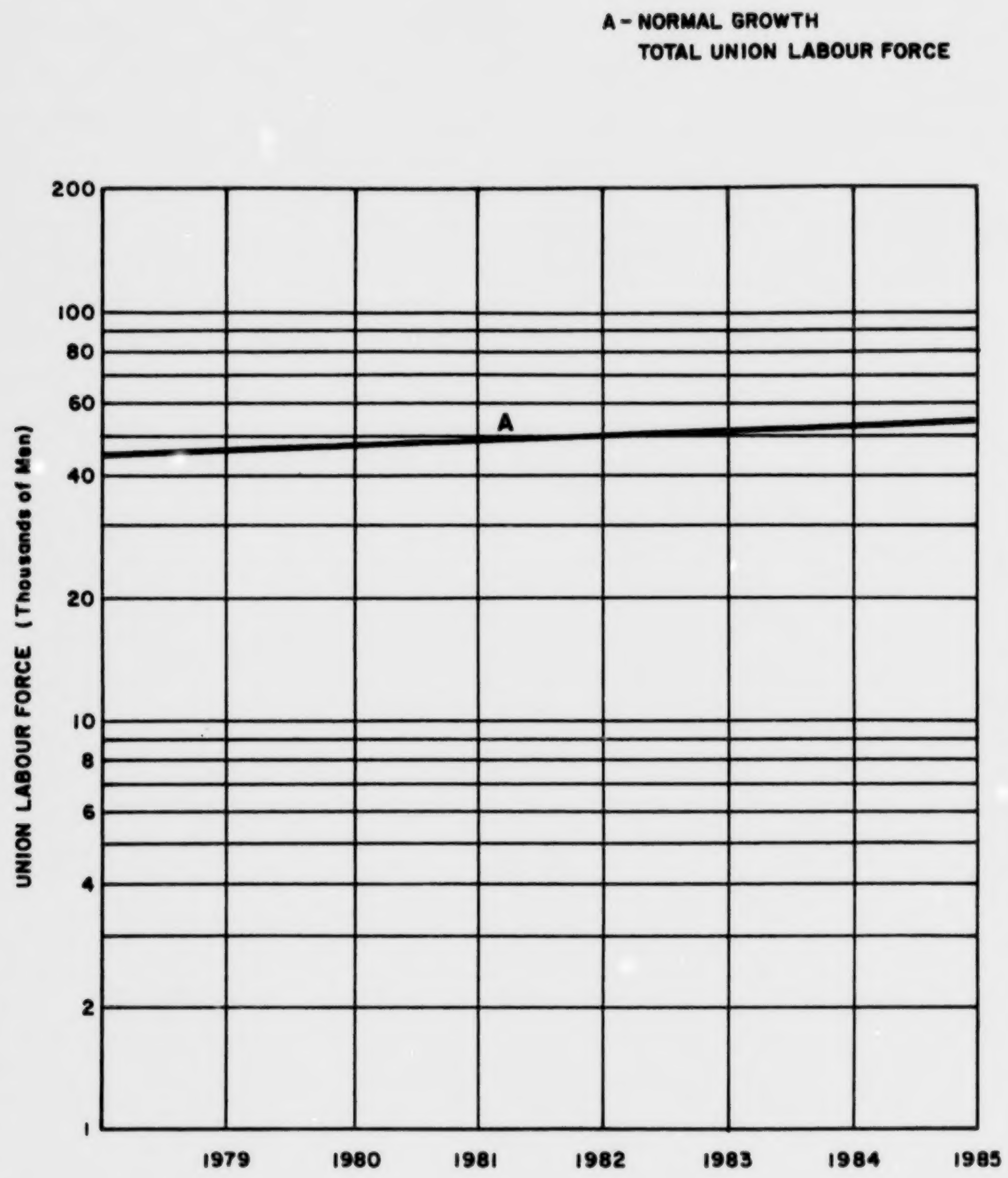
---

[1] "Big Inch Contractor", Oil Week, February 9, 1981
"Forecast", Wood Gundy, Canadian Economy

"Canadian Refining Industry To Spend $1.5 Billion On Refineries", Oil Week, February 9, 1981

"Petrochemical Industry To Make Considerable Investment in Alberta", Oil Week, February 9, 1981

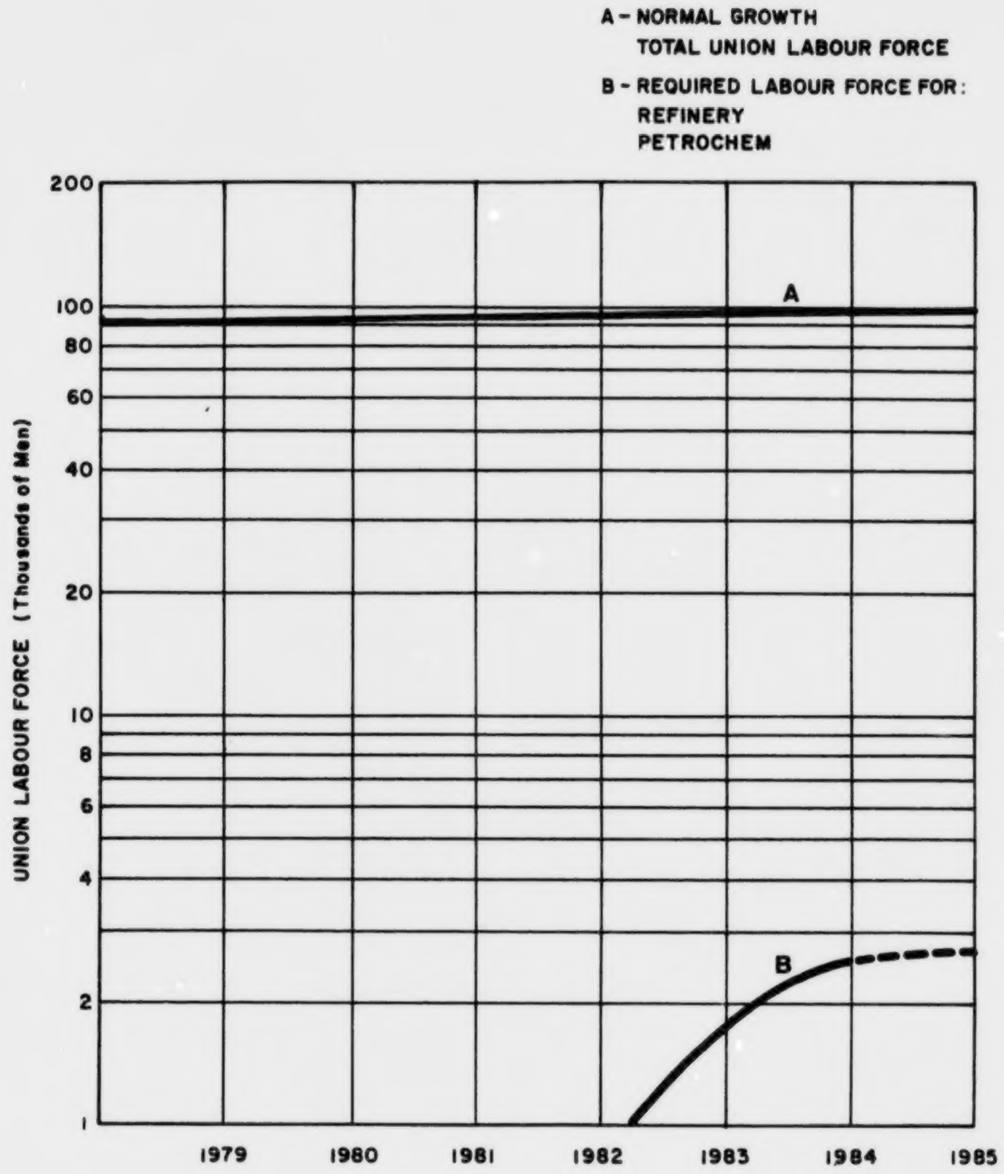
"Gas Plants Investment Boosted 50%", Oil Week, February 9, 1981



TOTAL MANPOWER DENSITY PROFILE

FOR THE ATLANTIC REGION

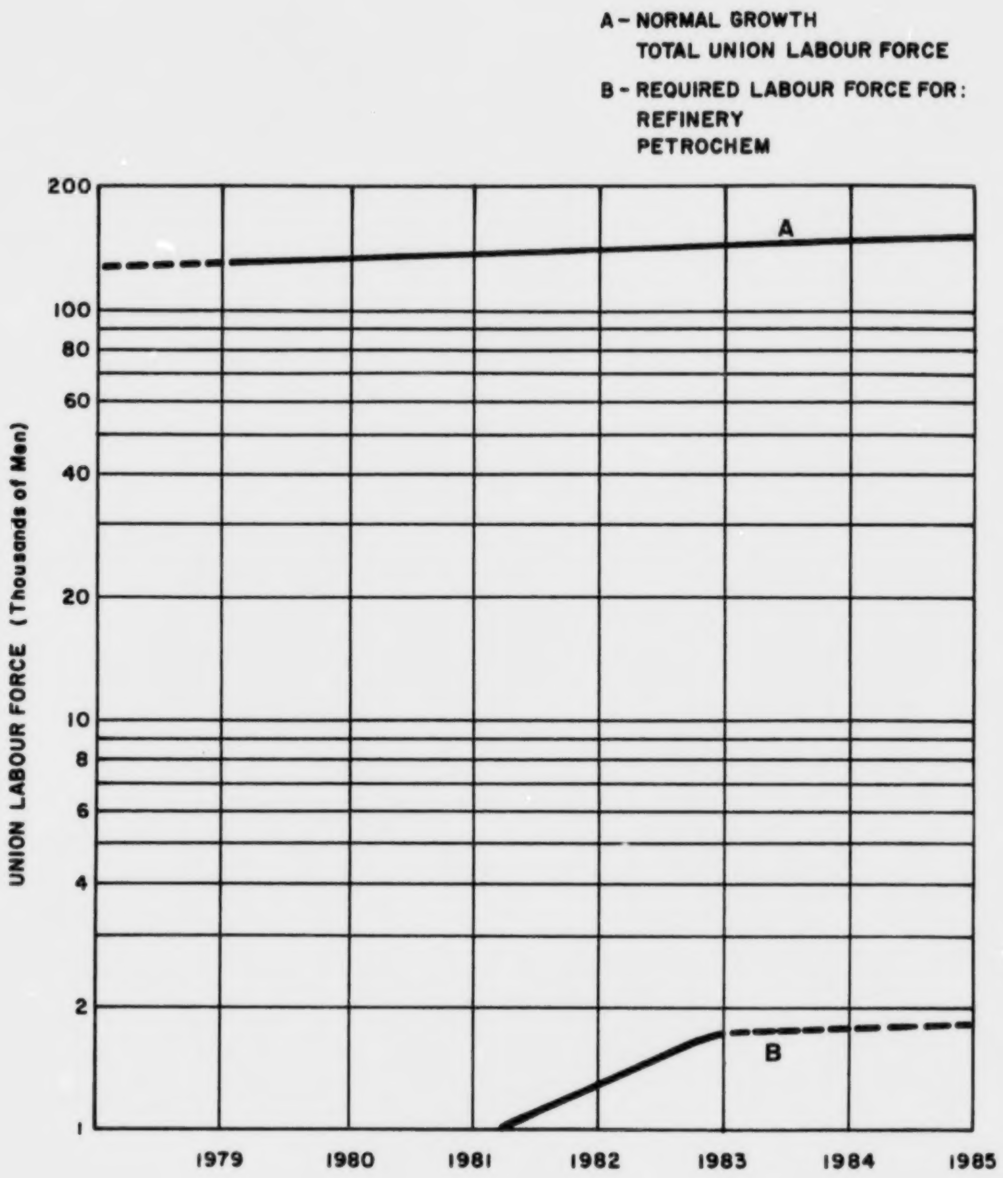
FIGURE 4·8



TOTAL MANPOWER DENSITY PROFILE

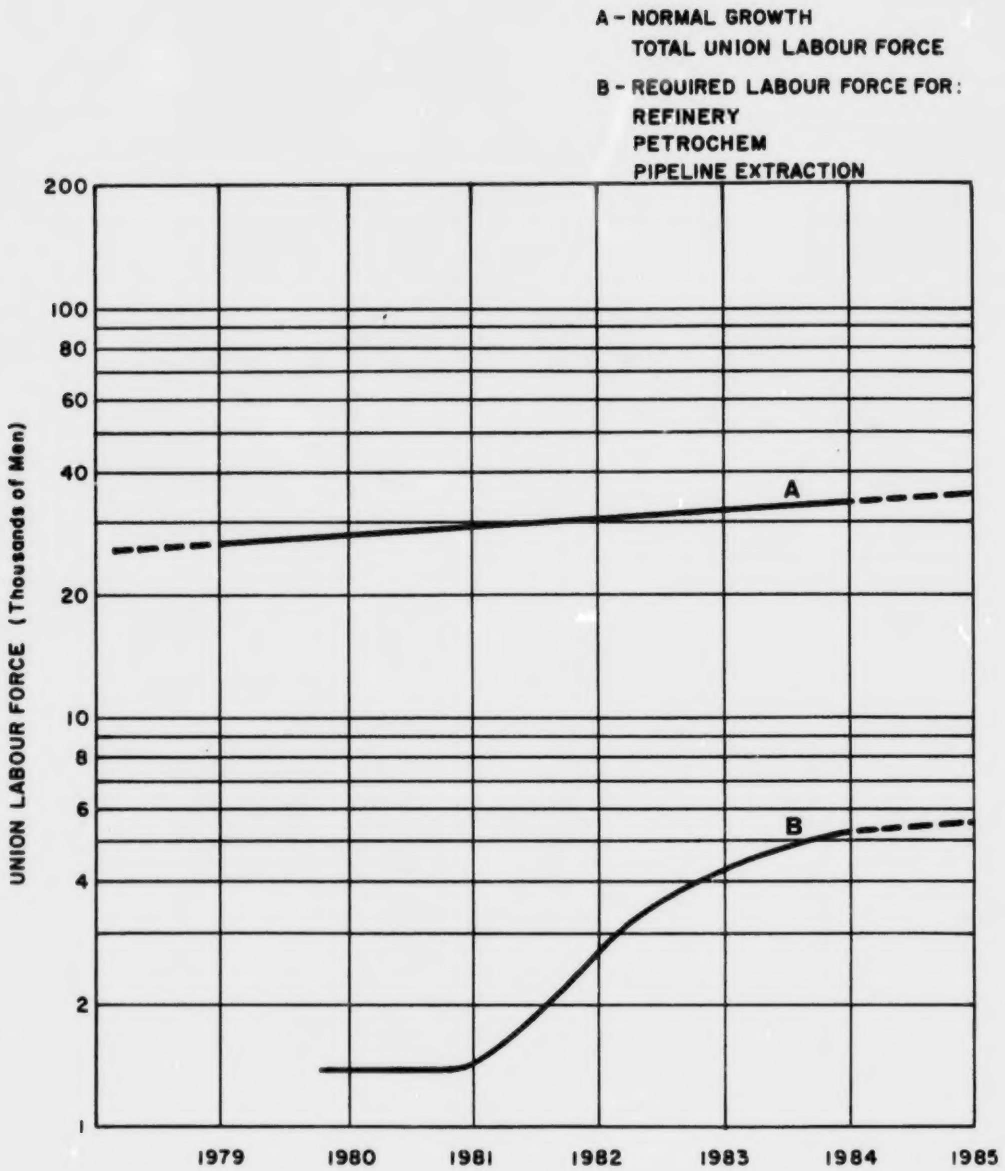
FOR THE QUEBEC REGION

FIGURE 4·9



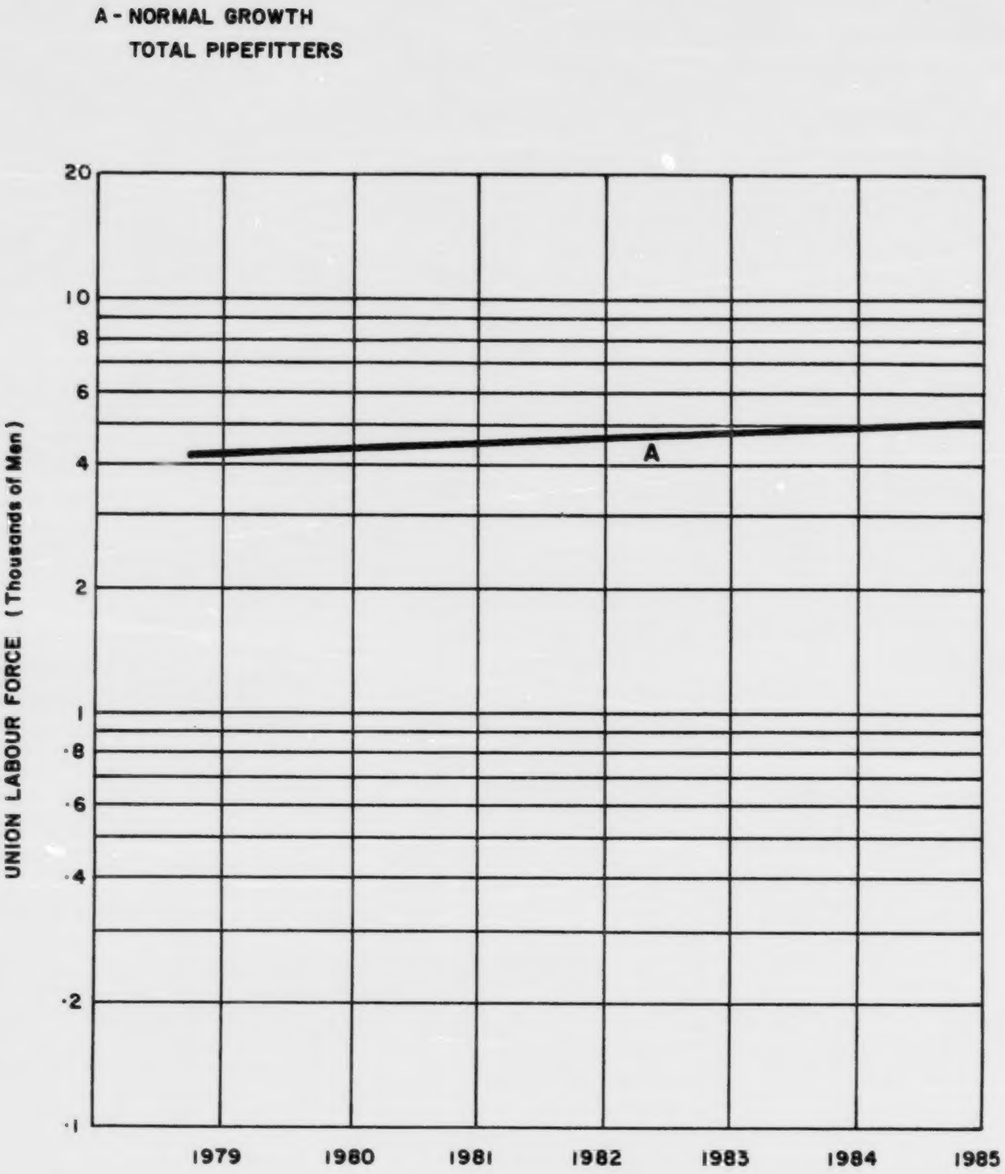
TOTAL MANPOWER DENSITY PROFILE

FOR THE ONTARIO REGION FIGURE 4·10



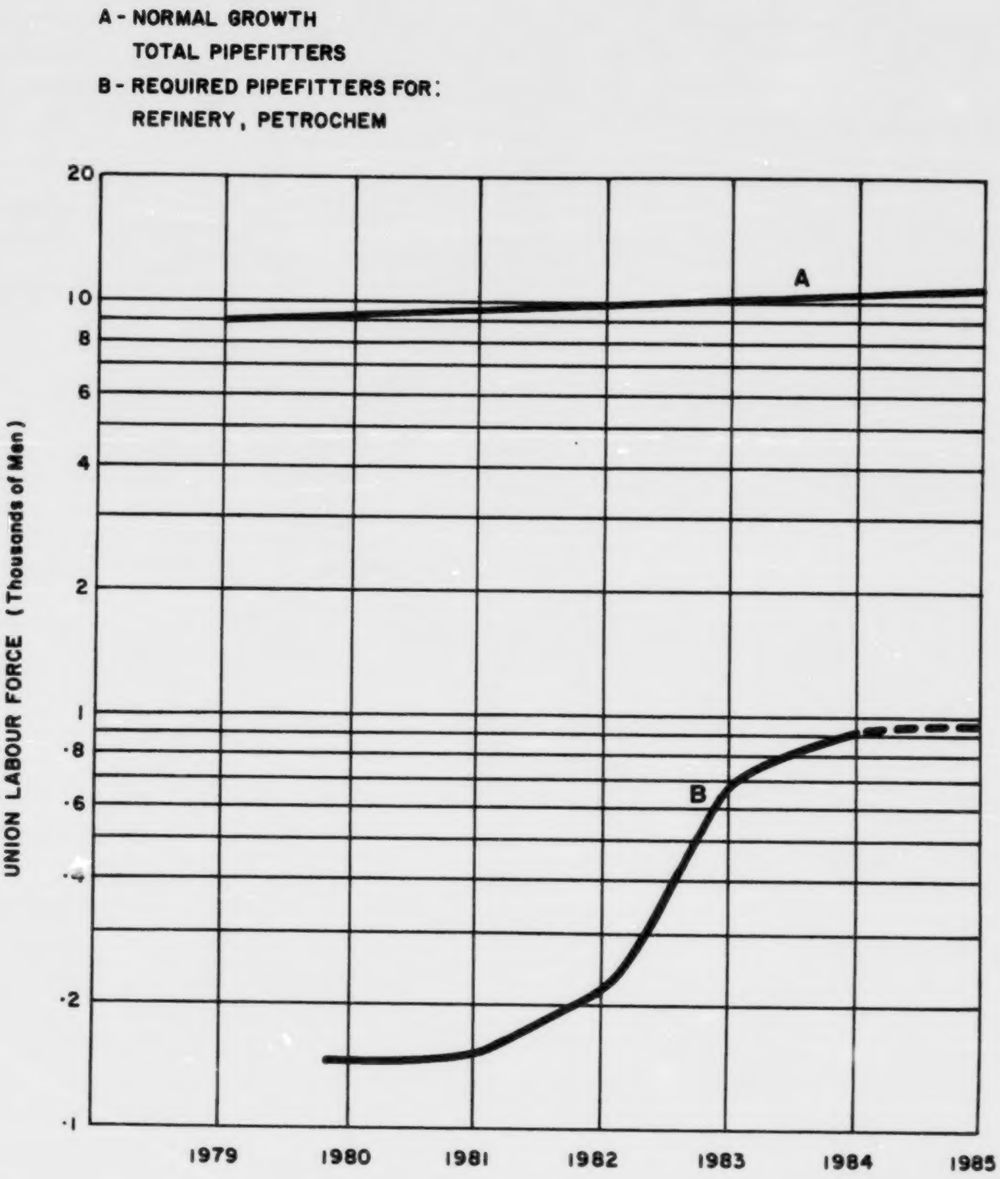
TOTAL MANPOWER DENSITY PROFILE

FOR THE WESTERN REGION FIGURE 4·11



PIPEFITTER MANPOWER DENSITY PROFILE FOR THE ATLANTIC REGION

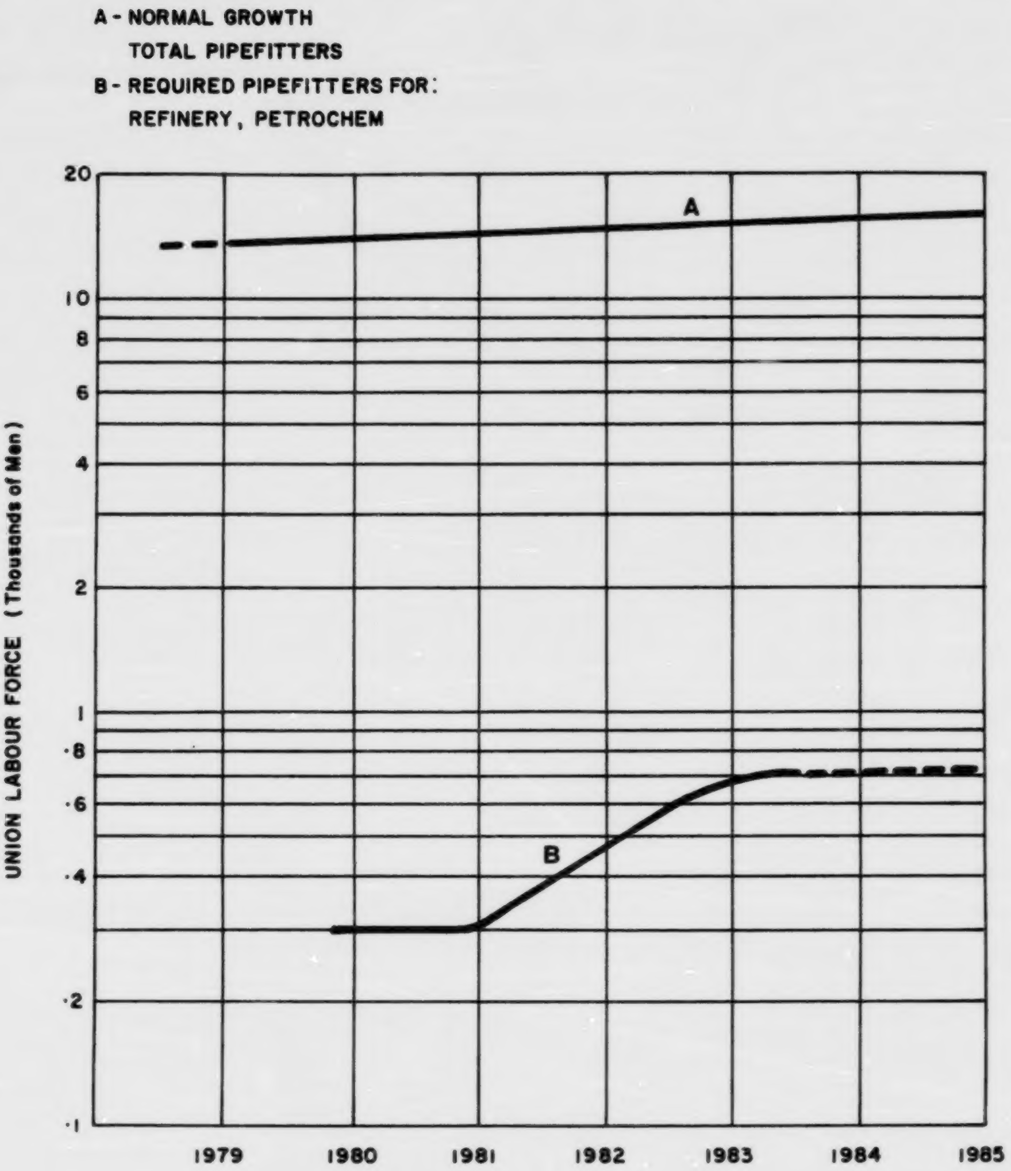
FIGURE 4·12



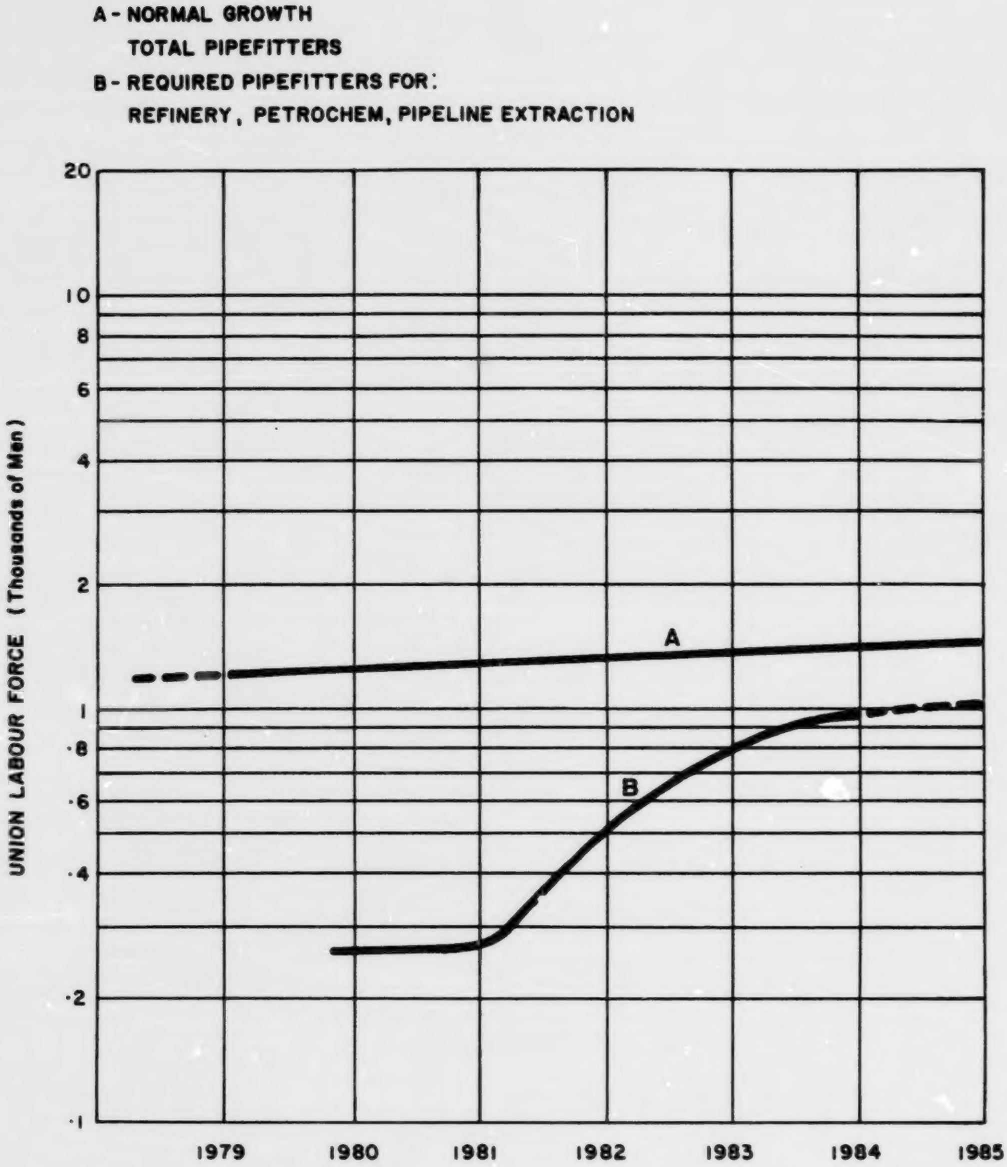
PIPEFITTER MANPOWER DENSITY PROFILE

FOR THE QUEBEC REGION

FIGURE 4·13



PIPEFITTER MANPOWER DENSITY PROFILE
FOR THE ONTARIO REGION FIGURE 4·14



PIPEFITTER MANPOWER DENSITY PROFILE

FOR THE WESTERN REGION FIGURE 4·15

TABLE 4-12

	YEAR	*ACTUAL MANPOWER[2] AVAILABILITY	*FORECASTED MANPOWER DENSITY NO. OF MEN	AS A % OF ACTUAL	% RISE IN DENSITY
Western	1980	142,000	7,300	5.1	0
	1981	151,000	13,572	9.0	86.0
	1982	158,000	21,837	13.8	199
	1983	163,000	26,263	16.1	260.0
	1984	173,000	28,000	16.2	283.6
	1985	180,000	28,000	15.6	283.6
Ontario	1980	136,000	834	0.6	0
	1981	138,000	1,270	0.9	52.3
	1982	141,000	1,710	1.3	105.0
	1983	143,000	1,820	1.3	118.2
	1984	148,000	1,890	1.3	126.6
	1985	153,000	1,890	1.2	126.6
Quebec	1980	92,000	400	0.4	0
	1981	93,000	573	0.6	43.3
	1982	94,000	1,795	1.9	348.8
	1983	95,000	2,424	2.6	506.0
	1984	96,000	2,630	2.7	557.5
	1985	97,000	2,630	2.7	557.5
Atlantic	1980	49,000	N / A**		
	1981	50,000			
	1982	51,000			
	1983	52,000			
	1984	53,000			
	1985	54,000			

* For basis see end of this section.

** No "planned" expansion of refineries and petrochem plants.

[2] "Corporations And Labour Unions Returns Act", Statistics Canada, Catalogue 71-202 and Catalogue 71-202S.

The petroleum industry will need more than its "fair share" of the total construction labour force. If the current situation can be considered a "fair share", the "peak" requirement of 16% of the labour market by 1983 in the Western region will cause a considerable amount of competition for labour, with respect to Canadian industry as a whole.

This situation can further be exemplified if we focus, for instance, on pipe fitters. The present "fair share" in the Western region is approximately 21% of the total construction labour force. Assuming that all major development projects do get underway, this percentage could increase to approximately 70%. Thus, an already potentially difficult situation becomes even more unlikely, when a labour demand of this magnitude can hypothetically be expected.

In addition to all aforementioned "fair shares", the effects of implementation of the Lead Reduction Programme must be considered. The "Engineering, Procurement and Construction Manpower Projection" schedule, included in this report, suggests the following critical peak manpower loadings:-

Western Region = 780 Men/Year
Ontario Region = 453 Men/Year
Quebec Region = 295 Men/Year
Atlantic Region = 300 Men/Year

Adding this potential loading onto the accelerating "Manpower Density" will, within the forecasted period, further aggravate Canada'a labour supply difficulties.

This can be illustrated using Table 4-12, which shows the Forecasted Manpower Density for the refining, petrochemical and related projects. If the increase in manpower density occurs during the critical peak years of 1984 - 1985 (see Figures 4-8 to 4-15), the combined effect on the manpower density would be the following:-

Western = From 16.2%* to 16.6%
Ontario = From 1.3%* to 1.6%
Quebec = From 2.7%* to 3.0%
Atlantic N / A.

* From Table 4-12.

4.0 COSTS

4.5.1 (Cont'd)

BASIS OF MANPOWER FORECAST

Forecasted Manpower Density

The total capital investment cost for all "planned" petrochemical, refining and related projects, from 1980 to 1985[1], is totalled and grouped into regions. Field construction manhours are determined from this total investment cost by means of factorial relationships obtained from construction experience from actual projects.

This direct cost factor relates labour, material and equipment components to a total cost structure of a complete refinery plant and can be assumed constant for a specific process. Variations due to productivity associated with the unique requirement of each refinery location are incorporated.

Using an "S" curve and a manpower histogram representing the most probable manpower density and field duration, an approximate distribution of the aforementioned costs is constructed. This, when converted, to manhours, results in a profile which represents the "forecasted labour force", shown on Figures 4-8 to 4-15.

Actual Manpower Availability

The total union manpower census in Canada, for all labour-related activities, is available, by regions, from Statistics Canada's Catalogue 71-202 and 71-202S, "Corporations And Labour Unions Returns Act".

[1] Refer to first page of Section 4.5.1.

CRAFT MIX

The following typical Craft Mix Relationships for total projects (as a per cent of total direct labour manhours) has been adopted.

Boilermaker	7%
Bricklayer	1%
Carpenter	14%
Cement Mason	2%
Electrician	9%
Insulators	2%
Ironworkers	8%
Millwrights	2%
Laborers	13%
Operating Eng.	4%
Painters	1%
Sheetmetal Workers	1%
Teamsters	1%
Piping	35%
	100%

4.5.2 Engineering and Home Office Manpower Forecast

In the same way that labour supply is expected to be a problem in the forecasted period, so will be the manpower required for engineering, procurement, field supervision and project management.

A 1980 survey[1] performed by ESSO Resources Canada Limited reports that consultants engaged in process plant design can accommodate an additional 32% technical work load. This is assumed to be representative of total engineering manpower availability, and can be used as a basis of comparison for the following "Projected Engineering Manpower Requirements".

Further, the Technical Service Council reports[2] 7,360 vacancies across Canada for accountants, engineers, scientists and other professionals. The report goes on to say that presently, "There is no general shortage of engineers. Rather, employers are seeking people with very specific training."

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[1] 1980 Canadian Process Plant Cost Survey, ESSO Resources Canada Limited.

[2] "Jobs For Professionals at Record Levels", Technical Service Council Report N15/80.

FORECASTED ENGINEERING MANPOWER DENSITY REQUIREMENTS FOR PLANNED PROJECTS (NOT INCL. LEAD REDUCTION PROGRAMME)

Year	Engineering Manpower Requirements (Men)	Percent Rise In Density
1980	5,470	-
1981	14,594	167
1982	16,381	200
1983	19,904	260
1984	19,000	250

The implementation of the Lead Reduction Programme will require the following Engineering Manpower in addition to the above.

Linear Time Scale (Year)	Engineering Manpower Requirements (Men)
1	375
2	521
3	574
4	413
5	52

4.0 COSTS

4.6 ESCALATION

With the use of the enclosed index, the cost in 1979 can be converted for another year based on the following formula.

$$\frac{\text{Price in Year}}{\text{Price in 1979}} = \frac{\text{Index for Year}}{\text{Index for 1979}}$$

Based on November, 1980 Wood Gundy's publication "Canadian Economy", and MHG resources, anticipated applicable escalation indices are shown below:-

Year	Index
1979	100
1980	112
1981	127
1982	147
1983	167
1984	188
1985	209

The three components of the cost structure, namely equipment, material and field labour, have been escalated by recommended percentage and weighted as per related cost mixes, to form the above cost factors. These have been applied to the total capital costs for conversion, pertaining to each refinery location.

4.0 COSTS

4.7 LEAD ANTI-KNOCK EQUIPMENT AT REFINERIES

Lead anti-knock additive is a toxic liquid and is usually stored at refineries in steel horizontal pressure vessels equipped with weigh scales and being nominally of 8000 gallon capacity. Only one tank is installed generally, but some refineries may have two depending upon geographic location and market demand.

Transfer of additive from tank car to refinery storage vessel is performed by inducing a vacuum on the latter using an inert gas (e.g. Nitrogen).

Transfer and mixing of additive with motor gasoline is effected by pumping the gasoline through an eductor, thereby inducing a controlled quantity of additive via a meter from the storage tank. Pumps handling the lead additive are thus obviated and steel lines used to transfer the fluid would be of small diameter only. Storage tanks and lines, etc., handling the leaded gasoline would be retained when the transition to a totally unleaded gasoline pool is effected. The ingrained lead in the steel will leach out slowly over an extended period, but overall the associated quantities of lead would be insignificant as to not be problematical regarding gasoline quality.

Equipment for storage of lead additive is usually housed in a separate building which would probably require the roof to be lifted or walls taken out in order to remove the tank(s) for decommissioning. This equipment could not be re-used and would have to be disposed of carefully.

Cutting up of the steel equipment is not allowed legally at the site due to the associated health hazard. Consequently, the equipment would have to be dismantled to a steel refinery for lead incineration and melting down.

Ethyl Canada estimate an expenditure of approximately $30,000 per refinery would be incurred for removal of equipment. This figure includes the transportation of the tank(s) to Baton Rouge (U.S.) for incineration.

ESTIMATE SUMMARY

NO LEAD

All $ x 1000

REGION	REFORMER CAP. COST	ISOMERIZATION CAP. COST	ALKYLATION CAP. COST	HYDRO TREATER CAP. COST	TOTAL CAP. COST
Atlantic	119600	16565	40890		177055
Quebec	150430	19195	30480		200105
Ontario	176755	30315	48777		255847
Western	121890	41320		90000	253210
TOTAL	568675	107395	120147	90000	886217

PAGES 132-167 HAVE BEEN DELETED TO PRESERVE CONFIDENTIALITY OF COST DATA FOR INDIVIDUAL REFINERIES

SECTION 5

5.0 GENERAL DISCUSSIONS

5.1 FUTURE GASOLINE PRODUCTION

5.1.1. FORECASTED DEMAND FOR GASOLINE TO 1995

5.1.1.1. General

Historically, motor gasoline has been the most important petroleum product in terms of volumes consumed. This is illustrated in Figure 5.1 in which it can be seen that in 1979 motor gasoline accounted for some 37% of the volume of all petroleum products consumed in Canada. The next most important product, heavy fuel oil, accounted for only 15.5% of the total consumption of petroleum products.

Of the motor gasoline consumed in Canada in 1979, some 93% was used in the transport sector of economy, the remainder being used largely in agricultural activities. Within the transport sector effectively 100% of the motor gasoline is used in the on-road transport mode with the automobile being the primary user. In addition, most small delivery trucks, pick-up trucks and vans run on motor gasoline as do a limited number of buses. Virtually, all large trucks and most buses operate on diesel fuel.

5.1.1.2. Growth in Demand for Motor Gasoline

The data of Table 5-1 indicates that the consumption of motor gasoline has increased in each year of the period 1965-1979. However, while the average annual rate of increase was close to 6% in the years prior to the formation of OPEC in late 1973, the average annual rate of increase has been 2.6% during the period 1973-1979.

5.1.1.3 Projections of Motor Gasoline Requirements

As motor gasoline has traditionally accounted for a major portion of the demand for petroleum products, many forecasters have devoted significant effort to the study of the demand for this product. As a result, a number of computer-based mathematical models have been developed. The following organizations are amongst those who maintain motor gasoline forecasting models and update the input data at regular intervals:

E.I.du Pont de Nemours (DuPont)
Ethyl Canada Inc. (Ethyl)
National Energy Board (NEB)
Energy, Mines and Resources, Canada (EMR)

In general these models estimate annual motor gasoline requirements over 10-20 year forecast periods on the basis of projections of the following:

1. The number of cars of each size and age which are in turn projected based on:
    - new car sales by size
    - scrappage rates for existing cars

2. Miles travelled per car per year which are in turn projected based on:
    - historic data on miles travelled per car per year
    - fuel prices
    - economic activity

3. Fuel economy which is in turn projected based on:
    - engine technology
    - user habits
    - legislation

In order to illustrate the interrelationships between these variables, historic and forecast data from a recent DuPont forecast are presented in Table 5-2. The forecast extends from 1979 to 1990 and, as a point of reference, historic data for the period 1968-1978 are included. It can be seen that DuPont's opinion is that the future will be quite different from the recent past. For instance, while the automobile population is shown to have increased at an average annual rate of 5.2% during the period 1968-1978, the forecast is for an average increase of only 2.2% during the period 1979-1990. Although complete details on DuPont's assumptions are not available, this might reflect expectations for slower population growth, restrained economic growth and increases in the real cost of owning and operating a car.

In the case of fleet miles per gallon performance, no significant change is shown for the period 1968-1978. The forecast, however, is for an average improvement of some 3.5% per annum during the period 1979 to 1990. This forecast improvement is likely based on the assumption that Canadian manufacturers will strive to meet the new car fleet mileage standards set by the U.S. Environmental Protection Agency. As well as the manufacturers improving the miles per gallon performance of the various sizes of automobiles the consumer will also have to want to purchase a mix of vehicles that will permit the fleet miles per gallon standard to be realized.

In order to achieve higher fleet miles per gallon figures it is expected that a significant change will take place in the size of the standard automobile engine. In this regard the assumptions as to engine size on which a recent Ethyl motor gasoline forecast was based are summarized in Table 5-3 . It can be seen that a very significant shift in the size of the engines purchased in new cars is forecast. For instance, while historic data for the year

1975 indicate that some 65% of new cars sold in Canada in that year were equipped with V-8 engines the forecast for 1985 is for only about 6% of new cars to have V-8 engines.

The remaining variable in Table 5-2 is miles driven per year. A complete set of reliable historic data are not available for this variable although the National Energy Board estimates that in 1976 the average automobile travelled 9400 miles.* The forecast assumes a 0.4% average annual rate of decline in miles driven. This might reflect assumptions such as gasoline prices rising in real terms thus discouraging unnecessary use of automobiles, the improvement of public transportation facilities in urban areas and the spread of the conservation ethic.

Motor gasoline consumption forecasts as prepared by each of four organizations mentioned above are summarized in Table 5-4 . The actual consumptions for 1978 and 1979 as shown in Table 5-1 can be seen to be quite close to those forecasted.

In comparing the forecasts, the initial observation is that, although the forecasts were carried out by distinct organizations over a time span of almost a year and a half, there is very little difference between these national level forecasts. Taking the DuPont of Canada forecast (the lowest) as the reference, calls for a 1990 motor gasoline requirement of only 4.6% greater.

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* Canadian Oil Supply and Requirements, National Energy Board, September 1978, Page 125. Although no source is mentioned in this report an earlier paper prepared by members of the National Energy Board's Economics Branch entitled "A Model for Forecasting Passenger Car Gasoline Demand" mentions an unpublished Environment Canada Study as the source for similar data.

The second notable feature of the forecasts is that in each case a decline in the absolute requirements for motor gasoline is forecast. The National Energy Board forecast, which calls for the sharpest drop, is for a 5.5% decline in the annual requirement in 1990 with respect to the starting point of 238 million barrels in 1978. This is a very significant change relative to the recent historic pattern. As can be seen in Table 5-1 the average growth rate in motor gasoline requirements during the period 1973-1979 was 2.6%. Simply applying this annual rate of increase to the 1978 consumption level would yield a forecast of 333 million barrels in 1990. This is 46% higher than the Energy, Mines and Resources forecast (the highest of the four) for 1990.

5.1.2 Projections of Demand by Grade

In order to satisfactorily model refinery operations it is necessary to breakdown motor gasoline requirements between the various grades of the fuel. At the present time the following distinct grades are in use:-

Leaded Regular
Leaded Premium
Unleaded Regular
Unleaded Premium

As illustrated in Table 5-3, very significant changes are expected in the type and size of engines which will be installed in the new cars of the future. These changes are primarily in response to emission and fuel economy standards. Corresponding changes in the relative requirements of the various grades of gasoline will result. In particular, there will be a decline in the requirements for leaded premium fuel. At the same time, emission control methodologies are also resulting in a shift to unleaded fuels. The overall result is that the future split between grades will be quite different from the historic division. This is illustrated in Table 5-5 which shows the historic split between the various grades at the national level for the year 1979 and available forecasts of the requirements for 1980, 1985 and 1990. In reviewing the table it is interesting to note the following:-

- The very significant relative decline in requirements for leaded fuels (76% of total in 1979 to 26% in 1990 in the DuPont forecast) and the significant relative increase in the requirements of unleaded fuels (24% of total in 1979 to 74% in 1990 in the DuPont forecasts).

- The two forecasts agree very closely with respect to premium fuels. However, with respect to regular fuels Ethyl forecasts 40% of the total requirements to be for leaded regular in 1990 as opposed to 26% by DuPont.

For purposes of this study breakdown by grade have been estimated by using the average of the DuPont and Ethyl forecasts.

Composite Forecast of Gasoline Demands

An updated composite forecast of gasoline demand is shown in Table 5-6 . The column for the future total gasoline was constructed on the basis of the average period-by-period growth rates for the motor gasoline forecasts given in Table 5-4 . The breakdown between leaded and unleaded gasoline for each year has been estimated by using the average of the DuPont and Ethyl forecasts as shown in Table 5-5 . For the year 1995, the same split between leaded and unleaded gasoline as for 1990 has been used.

The results of Table 5-6 are shown graphically in Figure 5-2 .

TABLE 5-1

CONSUMPTION OF MOTOR GASOLINE IN CANADA 1965 - 1979

Year	Consumption (millions of barrels)	Year Over Year Increase (%)	Average Annual Growth Rate (%)
1965	130.6	-	5.7
1966	138.1	5.7	
1967	145.2	5.2	
1968	154.8	6.6	
1969	160.7	3.8	
1970	169.2	5.3	
1971	176.0	4.0	
1972	188.3	7.0	
1973	204.0	8.4	
1974	211.4	3.6	2.6
1975	219.0	3.6	
1976	224.8	2.6	
1977	228.2	1.5	
1978	236.0	3.4	
1979	240.8	2.0	

TABLE 5-1 (Cont'd)

SOURCES:

1. Data for the years 1965 - 1969 obtained from Detailed Energy Supply and Demand in Canada 1958 - 1969, Statistics Canada 57-505, Occasional, (Secondary Energy Supply, Availability, line 21).

2. Data for the years 1970 - 1975 obtained from annual editions of Detailed Energy Supply and Demand in Canada, Statistics Canada Catalogue 57-207, Annuals, (Secondary Energy Supply, Availability, line 21).

3. Data for the years 1976 - 1977 obtained from the Quarterly Report on Energy Supply and Demand in Canada, Statistics Canada Catalogue 57-003, Quarterly.

4. Data for the years 1978 obtained from the Quarterly Report on Energy Supply and Demand in Canada, 1V Quarter 1978, Statistics Canada 57-003, Quarterly.

5. Data for the year 1979 obtained from Refined Petroleum Products, July 1980, Statistics Canada 45-004, Monthly.

TABLE 5-2

VARIABLES AFFECTING THE DEMAND FOR MOTOR GASOLINE BY AUTOMOBILES

A) Annual Data

	Automobile Population		Miles Travelled per Year per Automobile		Fleet Miles per Imperial Gallon	
	Millions of Automobiles	Year over Year % Change	Miles	Year over Year % Change	MPG	Year over Year % Change
Historical Period						
1968	6.16	-	See Note (2)		17.58	-
1969	6.43	*4.4			17.68	+0.6
1970	6.61	+2.8			17.80	+0.7
1971	6.96	+5.3			17.88	+0.4
1972	7.41	+6.5			17.88	0.0
1973	7.87	+6.2			17.79	-0.5
1974	8.47	+7.6			17.61	-1.0
1975	8.87	+4.7			17.47	-0.8
1976	9.34	+5.3			17.53	+0.3
1977	9.84	+5.4			17.64	+0.6
1978	10.27	+4.4			17.75	+0.6
Forecast Period						
1979	10.60	+3.2	9,506	-	17.96	+1.2
1980	10.88	+2.6	9,432	-0.8	18.28	+1.8
1981	11.16	+2.6	9,355	-0.8	18.80	+2.8
1982	11.43	+2.4	9,288	-0.7	19.50	+3.7
1983	11.73	+2.6	9,235	-0.6	20.39	+4.6
1984	12.02	+2.5	9,188	-0.5	21.37	+4.8
1985	12.30	+2.3	9,185	0.0	22.35	+4.6
1986	12.56	+2.1	9,177	0.0	23.28	+4.2
1987	12.80	+1.9	9,163	-0.2	24.17	+3.8
1988	13.03	+1.8	9,142	-0.2	24.97	+3.3
1989	13.24	+1.6	9,117	-0.3	25.68	+2.8
1990	13.44	+1.5	9,086	-0.3	26.28	+2.3

B) Growth Rates

		Average Annual Rate of Change - %		
Time Period		Automobile Population	Miles Travelled per Year per Automobile	Fleet Miles per Imperial Gallon
Historic:	1968-1978	+5.2	n/a	+0.1
Forecast:	1979-1990	+2.2	-0.4	+3.5

TABLE 5-2 (Contd')

Notes to Table

1) The arrangement of this table is intended to illustrate, in simplified form, the variables that are involved in the determination of motor gasoline demand. There is no implication that the DuPont model works in such a simplified manner.

2) These data are not available in reliable form. While it is possible to estimate automobile population from car registration data and fleet miles per imperial gallon from manufacturers data it is much more difficult to allocate actual gasoline sales between automobile and truck uses. It is therefore not possible to determine the average miles travelled per automobile by calculation.

Source: Data are taken from Estimated Highway Gasoline Consumption Forecast 1968 - 1990, E.I., Dupont de Nemours, Inc., September 1979.

TABLE 5-3

Canadian Car Sales by Engine Type

Year	% of Total				
	Large V-8(1)	Midsize V-8(2)	Small V-8(3)	6 Cyl	4 Cyl
Historic Period					
1970	16	32	12	18	22
1975	18	33	14	15	20
Forecast Period					
1980	8	14	23	30	25
1985	--	--	6	43	51
1990	--	--	1	40	59

NOTES:
(1) Over 360 CID
(2) 340 - 360 CID
(3) Under 340 CID

SOURCE: 1979 Canadian Automotive Trends Talk, Ethyl Corporation of Canada Limited.

TABLE 5-4

Forecasts of Motor Gasoline Requirements in Canada
(millions of barrels per year)

Year	National Energy Board	Energy, Mines and Resources	DuPont	Ethyl	Avg.
A) Forecasts					
1978	238	237	223	230	
1980	238	243	235	238	
1985	233	243	223	235	
1990	225	228	218	227	
1995	219	n.a.	n.a.	n.a.	
B) Average Annual Rate of Change - %					
Time Period					
1978-1980	0	+1.3	+2.7	+1.7	+1.43
1980-1985	-0.4	0	-1.1	-0.3	-0.36
1985-1990	-0.7	-1.3	-0.5	-0.7	-0.80
1990-1995	-0.5	n.a.	n.a.	n.a.	-0.5

Sources:

National Energy Board:	Canadian Oil Supply and Requirements, National Energy Board, September 1978.
Dupont:	Estimated Highway Gasoline Consumption Forecast 1968-1990, E.I. DuPont de Nemours, Inc., September 1979.
Ethyl:	Information prepared by Ethyl Corporation of Canada Limited in July 1978.
Energy, Mines and Resources:	Unpublished computer study dated December 12, 1979.

TABLE 5-5

Sales Demand for Motor Gasoline by Grade
(% of national total)

	History	Forecasts					
		DuPont			Ethyl		
Grade	1979	1980	1985	1990	1980	1985	1990
Leaded Regular	65	55	33	26	60	43	40
Leaded Premium	11	--	--	--	1	--	--
Unleaded Regular	24	33	50	53	27	37	40
Unleaded Premium	-(1)	12	17	21	12	20	20
	100	100	100	100	100	100	100

Note: (1) Included with unleaded regular.

Sources: History

1979: Refined Petroleum Products, Statistics Canada, Catalogue No. 45-004 Monthly, September 1980 issue.

Forecasts:

DuPont: Estimated Highway Gasoline Consumption Forecasts 1968-1990. E.I. DuPont de Nemours, Inc., September 1979.

Ethyl: Ethyl Canada Inc., August 1978.

TABLE 5-6

Demand for Leaded and Unleaded Gasoline
(millions of barrels per year)

Year	Leaded	Unleaded	Total
History			
1978	182.9	53.1	236.0
1979	163.3	77.5	240.8
Future			
1980	142	102	244
1985	91	149	240
1990	76	154	230
1995	74	150	224

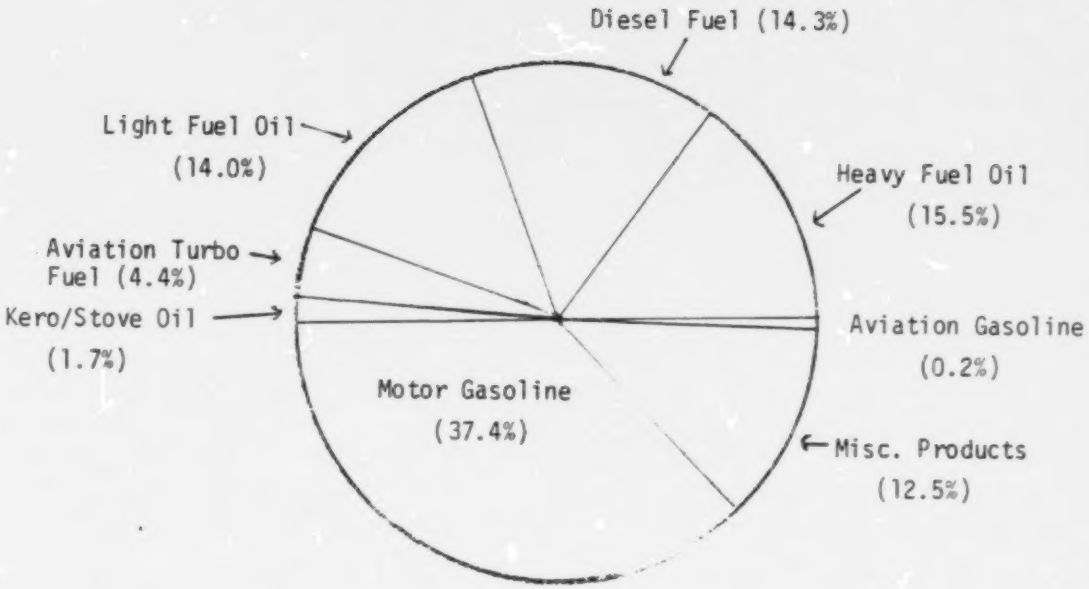

Fig. 5.1 DISTRIBUTION OF REFINED OIL PRODUCTS FOR 1979

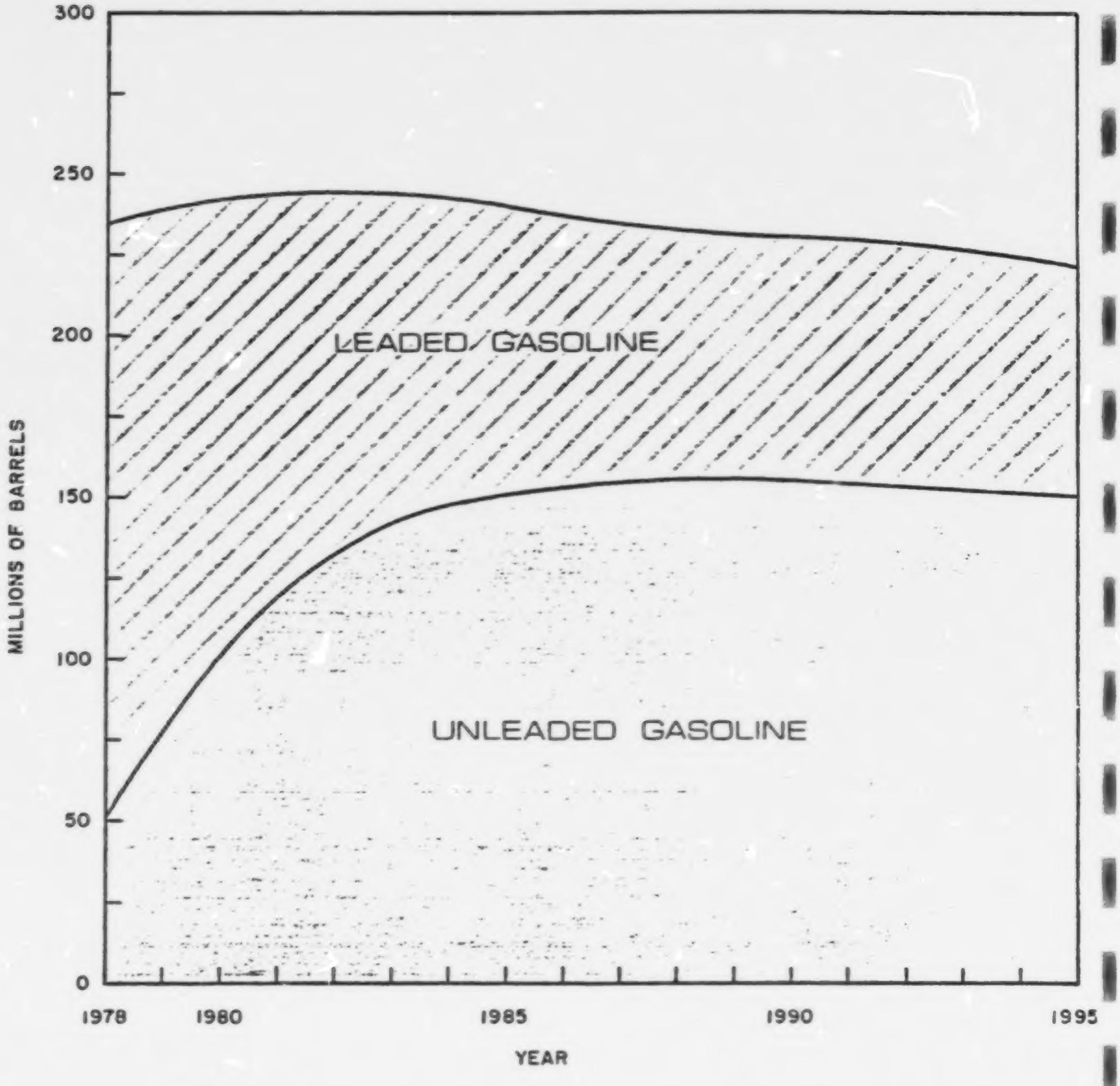

FIG. 5-2

FORECASTED DEMANDS FOR GASOLINE TO YEAR 1995

5.0 GENERAL DISCUSSIONS

5.2. REVIEW OF TECHNOLOGICAL IMPROVEMENTS TOWARDS LEAD-FREE GASOLINE PRODUCTION

Removal of lead anti-knock additives from gasoline drops its octane rating by some 5 to 20 numbers. There are basically two alternatives a refiner can use to compensate for this loss in octane quality. The first method is to alter processing routes by either installing new process units or modifying existing ones such that larger quantities of high octane streams are produced at the expense of lower octane streams. The second method is to substitute other antiknock additives. The various alternatives currently available within each of these two broader alternatives are discussed in some detail below.

5.2.1 Processing Improvements

Octane improving processes include catalytic reforming, catalytic cracking, alkylation, polymerization and isomerization. Current developments related to these processes can be broken down into two categories:-

a) Improvements resulting from catalyst developments and changes in operating variables within the units themselves

b) and improvements resulting from optimizing overall refinery schemes by integration of two or more processes.

Catalytic Cracking

Performance of existing units have been improved markedly in recent years. However, the improvements have been aimed particularly at throughput and gasoline yield increases rather than increases in RON. Increases in throughput by a factor of three have been reported while gasoline yields have been increased typically from 45% to 75% (Reference 17).

The major breakthrough leading to these improvements was the development of the active zeolite catalysts which began displacing the older alumina catalysts in the early sixties. Units adopting the new catalyst require physical modifications to convert from bed cracking to riser cracking which allows optimal control of the activity of the new catalyst. Additional improvements have been brought about by changes in operating variables such as increases in feed temperatures, increases in regeneration temperature and reduction of recycle slurry by installing more efficient reactor effluent cyclones.

As a result of increasing demand for lead-free gasoline along with lower forecast demand for total gasoline, the emphasis has recently shifted from yield and throughput improvements to selectivity improvements. Of particular importance is the development of catalysts providing higher gasoline RON. Recently, new catalysts have been developed which can increase RON by 2 - 3 points for the same operating conditions. (Reference 18.) This gain is significant when it is considered that cat cracker gasoline generally contributes 35% to 50% of the total gasoline pool. The new catalysts also strive to increase iso-butane and olefin production in order to provide additional alkylation feed.

Regarding operating conditions a trend towards higher severity (higher temperature and higher catalyst consumption) is envisaged in order to increase gasoline RON and olefin production at the expense of gasoline yield.

Catalytic Reforming

The approach used to improve the RON of reformate in recent years has been to strive for lower operating pressures without reducing catalyst life. Lower pressure increases the reforming reaction and hydrogen production but also favours coke deposition which fouls the catalyst. The majority of existing catalytic reforming units in Canadian refineries are of the semi-regenerative type.

Two major developments have been used to reduce the effects of catalyst fouling:

1. Continuous Regeneration. This technique was introduced about ten years ago. In this new design small portions of catalyst are removed from the last reactor bed, regenerated to a state equivalent to fresh catalyst and returned to the first catalyst bed. Other current processes employ cyclic regeneration by providing an additional valved reactor. This allows one reactor to be regenerated while the other reactors are on line. The older semi-regeneration processes which require periodical shutdowns for regeneration are becoming outmoded gradually because of the higher reforming severity required to achieve higher RON requirements.

2. Bimetallic Catalysts. These catalysts consisting of platinum and rhenium began displacing the older platinum on alumina catalyst in the late 1960's. These new catalysts provide yield and activity benefits as well as fouling resistance. New bimetallic catalysts have also been designed specifically for the older semi-regenerative, high pressure units mentioned above. This allows increases of throughput and/or severity for these units without reducing onstream time.

Isomerization

Two types of isomerization units are currently offered: C_5/C_6 units which convert normal paraffins to higher octane iso-paraffins and normal butane units which produce iso-butane as feedstock for alkylation units.

The C_5/C_6 units can provide an increase of 13 to 20 RON points depending on the extent of recycle of unconverted paraffins. To date, no isomerization units are known to operate in Canada while at least 30 units have been commissioned in the United States. As demonstrated in this study, the advantage of this unit is to allow an increase in octane barrels without increasing naphtha end points to levels that create problems in meeting distillate product specifications. The isomerization processing scheme is analogous to catalytic reforming in that the reaction takes place over catalyst in the presence of hydrogen. It is therefore possible to process the C_5/C_6 paraffins in the catalytic reformer, but this would result in lower overall yields and/or RON.

Isomerization/Catalytic Reforming Integration

As indicated above, the RON of isomerate can be increased by recycling unconverted normal C_5 and C_6 to the isomerization unit. In order to accomplish this, a separation unit such as molecular sieve adsorption is required. This adsorption unit could also be used to extract normal paraffins from the reformate, thereby increasing the RON of this component. The normal C_5/C_6 could be recycled to isomerization while the C_7+ normal paraffins could be selectively hydrocracked to produce iso-butane for alkylation. This scheme is commonly referred to as the aliphatic approach (Reference 19).

Alkylation/Polymerization Integration

Both these processes are used to convert olefins from the catalytic cracker into high octane gasoline components. Until the last few years, alkylation has been considered the preferred route because of its higher gasoline yield - one molecule of olefin produces one molecule of gasoline by alkylation, whereas two or more molecules of olefin are required to produce one molecule by polymerization. Two recent developments however, have created a resurgence of polymerization. These are the high cost of purchased iso-butane and the development of low cost polymerization techniques such as Dimersol. An approach which has recently been implemented, therefore, is to use both processes in tandem. (Reference 20). The objective of converting all of the olefins to gasoline is met by alkylating just enough olefins to consume all of the available iso-butane generated within the refinery and polymerizing the remaining olefins. Generally, schemes are devised to selectively polymerize the propylene and iso-butene olefins. All of 1-butene and 2-butene olefins which yield the highest RON alkylate are thereby dedicated to the alkylation unit.

Hydrocracking/Catalytic Cracking/Catalytic Reforming Integration

Hydrocracking can be introduced in a refinery scheme to alleviate refinery problems associated with increased diesel production as well as lead-free gasoline. For example, catalytic cracker cycle oil which is an undesirable diesel component because of its low cetane number, can be hydrocracked to produce high quality diesel and naphtha which can be used as reformer feed.

5.2.2. Alternate Antiknock Additives

Although many oil and chemical companies have searched for years for alternatives to lead antiknock additive, the only one that has found commercial acceptance is a compound of manganese (methyl cyclopentadienyl manganese tricarbonyl, or MMT for short) developed by Ethyl Corp. in the USA.

Depending upon the sources of the gasoline blending stocks and on its clear octane rating, 0.075 gm of manganese per imperial gallon of gasoline can raise the RON by up to two octane numbers. However, as a broad approximation, MMT is about twice as costly as TEL for the same octane number boost.

Some concerns have been expressed about possible health hazards with MMT, and that it might harm catalytic converters on car exhausts. The picture at the present is not clear with the result that the regulations governing the use of MMT are not uniform. For example, refineries in Ontario and Quebec are permitted to use up to 0.082 gm. of manganese per imperial gallon. However, the remaining Provinces have yet to specify an upper limit. In the USA, MMT can not be used in gasoline.

Other candidates for improving the octane rating of lead-free gasoline are not additives in the same context as TEL and MMT, but are actually high octane blending compounds used in proportions of up to say 15% of the gasoline blend. As such, they are better termed gasoline extenders rather than additives.

MTBE (methyl tertiary butyl ether) currently enjoys one of the fastest growing markets, particular ly in Europe, but also in the USA, providing petroleum refineries with an octane booster containing no metal as well as a means of extending the quantity of gasoline available. This results from its noteworthy octane blending performance. The range of MTBE blending octane numbers is:

RON	115-135
MON	98-110
(R + M) /2	106-122.5

The actual blending number generally rises as the gasoline base octane number decreases and as the gasoline base saturates content increases. Since MTBE has a low boiling point ($55^{o}C$), its addition improves the front end octane number. Engine performance of MTBE blended gasolines compared with all-hydrocarbon type gasolines show no difference in gasoline consumption or horsepower, sharp improvement in road octane number at low speed, no vapour lock or icing problems, no difference in evaporation losses and no corrosion problems.

MTBE is produced by reacting methanol with isobutylene. Currently all the MTBE manufactured in the USA to date is based upon the C_4 stream from ethylene plant steam cracking units which is a rich source of isobutylene. However, the total potential MTBE production associated with ethylene plants is small compared to the potential required for the unleaded gasoline pool.

A larger source for the required isobutylene is the fluid catalytic cracking unit in a refinery; however, MTBE will come in direct competition with the requirements for alkylation. Considerable study would have to be made on each refinery to determine which route, or whether a combination of the two, would be the most beneficial.

Methanol and ethanol are both very satisfactory fuels for internal combustion engines. They have high octane numbers and give low engine emissions, both properties being enhanced by the addition of a little water.

Some typical properties of methanol and ethanol are compared to those of a good quality premium grade gasoline in Table 5-7. It can be seen that the octane ratings of both alcohols are high, however per unit of volume, methanol has approximately 50% and ethanol 68% the energy of gasoline. Alcohol operated engines however, can be operated reliably at very lean fuel/air ratios, thus improving somewhat the foregoing percentages. Nevertheless, for a given travel radius, an automobile fuelled with methanol will require a fuel tank about 1.65 times as large as for gasoline. For ethanol as fuel, the figure would be 1.25 times. One particular problem of methanol usage is the cold start problem resulting from methanol's high heat of vapourization and the absence of light hydrocarbons.

Neither alcohol can be manufactured at a price anywhere near competitive to that for gasoline. This combined with the few minor engine modifications necessary has held back any extensive use of these fuels as an alternative to gasoline. Alcohol-gasoline mixtures have been tested extensively, and gasohol - a mixture of gasoline with 10% by volume of ethanol, is presently on the market in the USA. While the performance of gasohol in the engine apparently poses no problems, the production of the ethanol presents several financial and social questions. The cost of producing ethanol is somewhere in the neighbourhood of $2.00/gallon depending upon the starting materials, credits for by-products, sources of energy, etc. Thus, without some form of tax credit or price subsidy, gasohol's future as a fuel alternative is limited.

Methanol-hydrocarbon blends have a serious problem in that they exhibit a very low tolerance for water. Phase separation of water occurs at very low concentrations, and since it is practically impossible to exclude moisture from the vast gasoline marketing network, serious corrosion and freezing problems would result from its widespread use.

TABLE 5-7

PROPERTIES OF ALCOHOLS COMPARED TO GASOLINE

	Methanol	Ethanol	Premium Gasoline
Octane rating, unleaded:			
Research, R	112	110	96
Motor, M	92	90	84
(R+M)/2	102	100	90
Liquid density, lbs/imp. gal	7.96	7.94	7.44
Latent heat of vapourization at boiling point, BTU,/lb	473	368	370
Heat of conbustion, BTU/lb (liquid fuel to liquid water)	9776	13180	20260
Fuel/air, stoichiometric	0.155	0.111	0.069

5.0 GENERAL DISCUSSIONS

5.3 EFFECTS OF REFINING HEAVY CRUDES

5.3.1 Effects of Heavy Crude on Conventional Refineries

Yield structures for typical light and heavy crudes are compared below:-

		Domestic Light	Venezuelan Heavy	90% Light/ 10% Heavy
C4 and Lighter		5.5	1.0	5.05
Light Straight Run	(C5 to 200°F)	10.5	2.0	9.65
Naphtha	(200°F to 350°F)	17.0	6.0	15.9
Kerosine	(350°F to 500°F)	15.0	7.0	14.2
Light Gas Oil	(500°F to 650°F)	14.5	15.0	14.55
Heavy Gas Oil	(650°F to 1000°F)	24.5	32.0	25.25
Residuum	(1000°F to +)	13.0	37.0	15.4
Gravity (°API)		39.0	16.8	36.5

For illustration, a refinery may be required to modify its operation to process a blend of 10% heavy and 90% light crude as compared to 100% light crude and still meet the same product slate. The yield structure for this example is also included in the above table.

Due to counteracting factors, it is difficult to establish qualitatively what effect the new crude blend would have on the clear RON of the gasoline pool. The following factors are expected to influence the RON level:

1. The light straight run content of the gasoline pool would be reduced leading to higher octane level.

2. Reformer feedstock (naphtha) would be reduced leading to lower octane of the gasoline pool. As a result of the spare reformer capacity however, severity would be increased there by minimizing the loss of octane barrels.

3. The reduction in light straight run and reformate would be made up by additional cat cracker gasoline (crackate/alkylate/polymer). This additional gasoline make would result from an increase in the catalytic cracker feed that would be required to convert the higher yield of heavy gas oil. The increase in heavy gas oil yield would actually be higher than indicated in the above table because the boiling range of the heavy gas oil would have to be extended both at the front end and back end (say 640°F to 1010°F) in order to maintain product quality of distillates and heavy fuel oil.

   The RON of the cat cracker gasoline would be affected by two counteracting factors: conversion level required to satisfy the gasoline pool and quality of the feedstock. A higher severity and conversion level requirement is expected leading to high olefin production and higher crackate RON. On the other hand. the higher gravity of the feedstock would tend to lower the olefin yield.

In summary therefore, the gasoline pool composition would consist of a smaller fraction of light straight run, a smaller fraction of reformate having higher RON and a larger fraction of cat cracker gasoline with a different RON. The overall RON would therefore, be essentially unchanged.

5.3.2 Effects of Heavy Crude on Modifications for Lead-Free Gasoline Production

In the case studies presented in Section 3, the proposed approach is to up-grade the octane level of the light and heavy naphtha by installation of isomerization units as well as additional reformer capacity.

The potential of this approach is obviously reduced as the crude blends became heavier because of the lower availability of feedstock for these units. As discussed above, cat cracker gasolines become more predominant in the gasoline pool as the crude feed becomes heavier suggesting that cat cracker rather than reformer revamps would be favoured. The cat cracker revamps would entail modifications for increasing RON as well as debottlenecking the units to accommodate the higher yields of heavy gas oil. Possible modifications leading to higher octane would include conversion to riser cracking or to a high olefin type of operation in conjunction with expansion of the alkylation unit. In addition to cat cracker expansions, isomerization units would likely still be required. However, these would be smaller than those required for a light crude.

5.3.3. Heavy Oil Upgrading

The above discussions were based on a modest increase in heavy crude feedstock which could be handled by conventional refineries. However, as the proportion of heavy crude oil increases (or demand of light and heavy fuel oil decreases) a point will be reached where residual oil could no longer be absorbed by the product slate. In this situation heavy oil upgrading schemes would have to be introduced to convert the residual oils to gasoline and distillate components.

Flexicoking, hydrocracking, heavy oil catalytic cracking and visbreaking are four upgrading processes which can be used to reduce heavy fuel oil production without making byproducts. Delayed coking or solvent deasphalting are two other schemes which can be implemented depending on market demand for coke or asphalt. All these processes produce naphtha streams which can be fed to the reformer. This feature would tend to increase octane level without requiring expansion of the catalytic reformer which would have spare capacity due to the low yield of virgin naphtha.

5.0 DISCUSSIONS

5.4 OPPORTUNITIES FOR SPECIALIZATION

The capital cost estimate for the lead-free studies were based on the assumption that all refineries would be modified to include isomerization unit installations and catalytic reformer expansions. The cost of the unit installations and catalytic reformer expansions could be reduced by providing the same additional capacity in a reduced number of reformer expansions, thereby lowering the cost per barrel of installed reformer capacity. The number of isomerization units could also be reduced by installing units with recycling of unconverted normal pentane and normal hexane . This recycle feature typically increases RON of the isomerate from 83 to 89. Another step to reduce the number of isomerization units could be to increase the octane level of cat cracker gasoline as discussed in Section 5.3. As pointed out, this type of modification would be more appropriate for heavier crude diets.

From these considerations, two types of refineries are suggested each of which would specialize in different product slates:

1. Premium Gasoline Producers

   These refineries would install isomerization units with recycle and expand reformer capacity. In order to maximize feed to these new units, the crude diets would consist of light crude and condensate. Because of the high reformer feed rates, the naphtha end point would tend to be high. This would provide a relatively heavy kerosine in lower quantities thereby limiting blending flexibility in the distillate pool. Consequently, the diesel to light fuel oil ratio for these refineries would tend to be low.

2. High Diesel Producers

These refineries would have revamped catalytic cracker units which would allow them to process the heavier crudes more efficiently. These refineries could also reduce reformer throughput allowing them to increase the kerosine yield for high diesel and jet fuel production. Due to low reformate yields as well as lack of isomerate, these refineries would produce predominantly lead-free regular gasoline.

This type of optimizing scheme involving more than one refinery would be applicable only in areas where neighbouring refineries share a marketing region. Isolated refineries supplying local markets would obviously require total revamps, as suggested in the case studies in order to meet the lead-free requirement.

SECTION 6

6.0 CONCLUSIONS

6.1 It was decided to segregate Canada into four distinct regions, namely Atlantic, Quebec, Ontario and Western. These geographic regions are identified in Figure 2-3 . The rationale was to reflect the widely differing process parameters prevailing with respect to crude oil diet and product slate in 1979. These fundamental differences are illustrated in Table 2-20. There were a total of thirty seven (37) oil refineries operating in Canada in 1979 and the regional split is given below:

Atlantic	5;	Quebec	7
Ontario	8;	Western	17

Point Tupper refinery in the Atlantic Region closed down in early 1980 but has been included in this study for completeness. The Come-by-Chance Refinery was excluded however.

6.2 Lead additive was used in all regions to produce leaded regular gasoline. Gasoline pools were represented by three basic grades of differing octane quality and relevant 1979 data is given below:

Grade	Atlantic	Lead Level gms Pb/I.G. Quebec	Ontario	Western
Leaded Regular (93.1 RON)	2.71	2.76	1.73	1.99
Unleaded Regular (93.6 RON)	0	0	0	0
Unleaded Premium (97.5 RON)	0	0	0	0

6.3 The results of the process simulator indicate that no capital investment would be required by refineries within any region if lead level controls of 2 gm/IG and 1 gm/IG in motor gasoline were imposed. This conclusion is hypothetical and does not preclude the possibility of certain refineries requiring extra process unit capacity created by divergences in refinery process configuration, crude diet, product slate (+ seasonal product demand changes), from the regional Base Case. A capital investment figure of between $50M to $100M is provided as a best guesstimate for the 1gm/IG control.

6.4 The imposition of a zero lead level control definitely will create a need for significant capital investment by the refineries. The simulation results reveal that all regions require extra processing capacity as summarized in Table 4-1. Based on specific approaches adopted in the study, the types of process units required are confined to Naphtha Catalytic Reforming, LSR Isomerization (plus associated LSR Hydrodesulphurization units) and HF Alkylation. It has been deduced within the limitations of the study, that all other types of existing process units are adequate to cope with the extra demand of octane barrels to produce total lead-free gasoline pools. These units include Crude Distillation, Vacuum Distillation, Fluid Catalytic Crackers, and Catalytic Polymerization. It is pertinent to mention that the existing total capacity of Naphtha HDS units associated with the existing Reformers can marginally handle the extra naphtha demand, with the exception of the Western region which is deficient.

6.5 For all study cases pertaining to the 2 gm/IG and 1 gm/IG lead level controls, distillate physical property specifications could be met with relative ease, thus obviating the necessity to increase reformer severity above 100 RON. However, for the zero lead level cases, increased severity to 103 RON, and extra alkylation capacity is necessary in all regions except Western. Refer to Tables 3-2 to 3-5.

6.6 Incremental crude oil throughputs reach a maximum for the zero lead level cases as would be expected, and range from 0.47% to 2.26% across the four regions. Associated energy penalties increase generally in accordance with the reducing lead levels for all regions from a maximum of 0.18% at the 2 gm/IG control, 0.86% at the 1 gm/IG control, to a maximum of 0.82% at the zero lead control. Exceptions occur in Quebec and Atlantic regions and the anomalies are created by the masking effect of the MMT addition, which reduces the octane barrel demand.

6.7 The necessity to increase HFO production in order to meet product specifications has been obviated by raising reformer severities to 103 RON. Cetane index specifications for diesel fuels will become increasingly constraining as the percentage of syncrude in the crude diets rises.

Regional product demand ratios vary considerably and should be assessed when interpreting the results of this report. (Compare Tables 6-1 to 6-4).

With increasing control standards on the use of lead additives, it should be appreciated that the bulk of the net octane barrel demand is derived from high octane catalytic reformate. The higher the severity of operation of the reformers, the greater is the proportion of naphtha feedstock which is converted to fuel gas.

It is evident from the simulations that there is a trend of increasing availability of fuel gas surplus to process unit requirements, as the lead levels are reduced to zero.

Table 6-5 reflects this trend.

It is important to note that the fuel gas requirements in Table 6-5 do not include offsite fuel such as that used for steam raising, etc. which is a big proportion of the total refinery fuel consumption (normally 7% of crude). Consequently, this theoretical surplus of fuel gas could in practice back-out the HFO used as refinery fuel, to make more HFO available for sales.

However, it may be more profitable for the refiners to sell LPG instead.

6.8 Order of magnitude capital investment estimates (1979) were developed for the zero lead cases corresponding to extra process unit capacities required and they are summarized below:-

	$ million Canadian (1979)
Atlantic	177
Quebec	200
Ontario	256
Western	253
TOTAL	886

These costs reflect allowances for regional productivity variations, normal construction labour availability, smallness of incremental process capacity required, and offsites expansion at the refineries.

6.9 Regional availability of construction labor has been assessed and graphically represented in Section 4. Published data on future refinery and petrochemical projects shows that skilled trade construction labor availability will be a major constraint dictating their go/no go decisions. Pipefitters are forecasted to be the key constraint, especially in the Western region.

The engineering, procurement and construction manpower loads created by the zero lead program will aggravate this situation. It is proposed that these refinery revamps and expansion be staggered on a regional basis and the associated construction schedules developed for this report have been similarly based. A four to five year program is considered to be realistic, but its feasibility will depend on extraneous factors such as natural manpower shifts within Canadian provinces and to the USA, in addition to probable extended equipment/material delivery periods. It is envisaged that project deferments will be necessary in the Western region.

6.10 Due to the expected non-conformity of many refineries with the composite refinery model, it is envisaged that most refineries will have to undergo modifications in one form or another.

The number of refinery employees will have to increase to ensure adequate control and operation of the extra process units for the lead case. Based on the 'average' refinery, it has been estimated that an additional 900 to 1000 refinery employees will be needed within specific classifications.

6.11 Incremental process operating costs compared to the Base Case have been calculated for all study cases, and the overall results are shown below. Allowances have been made for regional crude oil, i-C_4 and HFO prices, wages, reduced lead additive usage, catalyst replacement and refinery fuel consumption.

$1000 Can/CD

Control Case

Region	2 Lead	1 Lead	No Lead
Atlantic	16	35	43
Quebec	23	94	67
Ontario	0	7	60
Western	0	1	41

Reference should be made to Table 2-17 when interpreting these results.

6.12 Increased inventory costs have been interpreted as extra crude oil storage, necessarily induced by the incremental crude oil throughput. Based on an assumed strategic requirement of 30 days storage, this translates to a 'one-off' extra crude oil cost.

Working capital includes return on capital investment (25%), insurance, local taxes, and maintenance costs. See Table 4-10 for details.

In summary, the total incremental cost allowing for all of the above factors, to produce motor gasoline containing reduced lead additive levels are shown below. The cost can be based on either the total motor gasoline pool volume or the converted regular leaded portion only. The figures below pertain to the converted regular leaded portion of motor gasoline pool.

		Canadian cents/litre	
		Control Case	
Region	2 Lead	1 Lead	No Lead
Atlantic	0.20	0.43	2.52
Quebec	0.17	0.69	1.80
Ontario	-	0.03	1.37
Western	-	0	1.05

6.13 Obsolete refinery equipment caused by the reducing lead level control program, is difficult to assess. The approach adopted in this study was one whereby all existing process units were utilized to the fullest extent possible.

However, due to circumstances specific to individual refineries, it is envisaged that some process units may become obsolete but cannot be identified at this stage. Some old catalytic reformer units of low capacity may be too expensive to modify to high severity operation and thus may be scrapped and replaced completely by a new unit of larger capacity based on latest technology.

It should be noted that if the zero lead control is imposed, the lead additive storage and injection control equipment will become redundant. Dismantling procedures are very restrictive and equipment would have to be sectionalized and transported to a suitable steel refinery for incineration. Costs for this work have been estimated to be $30,000 per refinery.

6.14 The future demand for gasolines to the year 1995 has been forecasted. Total demand for motor gasoline is expected to decline slowly beginning 1982 with the percentage of unleaded grades increasing rapidly until 1987 when a levelling-off trend is envisaged. The above assumes that lead-level controls are not legislated.

6.15 It is apparent from the geographic locations of Canadian refineries that there are definite opportunities for product specialization and are categorized as 'premium gasoline' and 'diesel fuel' type refineries. Remote location refineries will incur the greatest expenditure to produce lead-free gasolines. Heavy crude refining seems appropriate for all regions to supplement the shortage of middle distillate fuels in the Western region and to convert the surplus heavy fuel oils to more saleable products in the Eastern regions. Hydrocracked gasoline has a relatively low octane quality but can be upgraded by reforming.

Alternative processing schemes are discussed briefly in the report including alternates to lead anti-knock additives such as lead-free additives (MTBE) and alcohol.

6.16 It should be appreciated that many refineries will deviate from the generalized model of composite refineries used in the study. The capability of existing catalytic reformers and catalytic cracking units (FCC) to accommodate high throughputs at high severity/conversion levels cannot be assessed and is sure to differ widely from one refinery to another. Additionally, the volume percentages of virgin naphtha (LSR) which are currently used for the gasoline pool are not known. The refinery simulations have revealed that this figure will have a major influence on the relative difficulties refineries will experience, to produce lead-free gasolines.

Only one key approach has been evaluated in the study. namely to upgrade the octane number of LSR. Other approaches which have not been considered are listed below:-

- catalytic cracking to raise octane numbers
- heavy oil upgrading to provide reformer feed
- alcohol addition, etc..

6.17 In conclusion, the study provides insight into some of the problems which can be expected and highlights directional trends which refiners might assess to effect a totally lead-free motor gasoline pool.

TABLE 6-1

Production of petroleum products - 1979 for the Atlantic Region

	Barrels Produced	Percentage of Total
Motor gasoline	27,575,181	25.76%
Aviation gasoline	0	0
Aviation Turbo Fuel	3,426,921	3.20
Diesel Fuel	9,913,121	9.26
Kerosene & Stove Oil	3,604,770	3.37
Light Fuel Oil	26,427,789	24.68
Heavy Fuel Oil	26,118,520	33.73
Total Transportation & Heating Fuels	107,066,302	100.00

gasoline/distillate ratio, defined as:

$$\frac{\text{motor gasoline \& aviation gasoline}}{\text{turbo fuel + diesel fuel + kero + stove + light fuel oil}}$$

$= 0.636$

Source: Petroleum Refineries in Canada, December 1979
Energy, Mines and Resources, Canada.

TABLE 6-2

Production of petroleum products (1979) for Quebec

	Barrels Produced	Percentage of Total
Motor gasoline	57,279,571	36.21%
Aviation gasoline	624,995	0.40
Aviation Turbo Fuel	7,885,300	4.98
Diesel Fuel	14,315,355	9.05
Kerosene & Stove Oil	6,446,064	4.07
Light Fuel Oil	34,375,055	21.72
Heavy Fuel Oil	37,302,020	23.57
Total Transportation & Heating Fuels	158,246,360	100.00

gasoline/distillate ratio, defined as:

$$\frac{\text{motor gasoline + aviation gasoline}}{\text{turbo fuel + diesel fuel + kero + stove + light fuel oil}}$$

$= 0.919$

Source: Petroleum Refineries in Canada, December 1979
Energy, Mines and Resouces, Canada

TABLE 6-3

Production of petroleum products (1979) for Ontario

	Barrels Produced	Percentage of Total
Motor gasoline	72,309,238	41.29
Aviation gasoline	79,705	0.05
Aviation Turbo Fuel	8,061,834	4.60
Diesel Fuel	18,800,409	10.74
Kerosene & Stove Oil	8,113,045	4.63
Light Fuel Oil	29,641,105	16.92
Heavy Fuel Oil	38,133,113	21.77
Total Transportation & Heating Fuels	175,138,449	100.00

gasoline/distillate ratio, defined as:

$$\frac{\text{motor gasoline + aviation gasoline}}{\text{turbo fuel + diesel fuel + kero + stove + light fuel oil}}$$

= 1.120

Source: Petroleum Refineries in Canada, December 1979
Energy, Mines and Resources, Canada

TABLE 6-4

Production of petroleum products (1979) for Western Region

	Barrels produced	Percentage of Total
Motor gasoline	83,595,601	49.76%
Aviation gasoline	867,818	0.52
Aviation Turbo Fuel	11,521,704	6.86
Diesel Fuel	50,223,113	29.89
Kerosene & Stove Oil	6,845,843	4.08
Light Fuel oil	1,090,058	0.65
Heavy Fuel oil	13,851,771	8.24
Total Transportation & Heating Fuels	167,995,908	100.00

gasoline/distillate ratio, defined as:

$$\frac{\text{motor gasoline + aviation gasoline}}{\text{turbo fuel + diesel fuel + kero + stove + light fuel oil}}$$

$$= 1.212$$

Source: Petroleum Refineries in Canada, December 1979
Energy, Mines and Resources, Canada

TABLE 6-5

REGIONAL REFINERY FUEL GAS BALANCES

	REFINERY OWN FUEL CONSUMPTION CALC'D FROM STATSCAN DATA*	CALCULATED EQUIVALENT REFINERY FUEL DEMAND FROM PROCESS UNITS (NOTE 1)	EQUIVALENT FUEL DEMAND FROM OFFSITES	ACTUAL FUEL GAS PRODUCTION FROM SIMULATION	LPG SALES	REFINERY FUEL CONSUMED	
						FUEL GAS	FUEL OIL
	BPCD	BPCD FOE	BPCD	BPCD FOE	BPCD	BPCD FOE	BPCD
BASE CASE							
ATL	15985	11696	4289	10556	1273	9283	6702
QUE	32829	22714	10115	20300	3586	16714	16115
ONT	49617	27133	22484	27454	7473	19981	29636
WEST	32464	25971	6493	25120	5346	19774	12690
2 LEAD							
ATL		12083	4289	11908	1273	10635	5737
QUE		24393	10115	23244	3586	19658	13950
1 LEAD							
ATL		12623	4289	12988	1273	11715	5737
QUE		24284	10115	26672	3586	23086	11313
ONT		28581	22484	33408	7473	25935	25130
WEST		27904	6493	27799	5346	22453	11944
NO LEAD							
ATL		12934	4289	14372	1273	13099	4124
QUE		25064	10115	27756	3586	24170	11009
ONT		30760	22484	38062	7473	30589	22655
WEST		30606	6493	35479	5346	30133	6966

Note 1:
Excludes all offsite refinery fuel consumption

* Statscan reports 20227

FUEL GAS

QUEBEC

* STATSCAN DATA

* Total own consumption = 1927000 M^3/yr
* Total crude output = 9806000 M^3/yr

∴ Fuel consumption = 6.5% of crude T/P

Base Case crude rate from simulation = 505062 BPCD

∴ Own fuel consumption = 505062 x 6.5% = 32829 BPCD

Base Case fuel gas production = 20300 BPCD FOE

FROM STATSCAN DATA :

LPG SALES = 3586 BPCD FOE

∴ Fuel gas available = 20300 - 3586 = 16714 BPCD FOE

Total demand =32829 BPCD
Fuel Gas supplied =16714 = 50.9%
Fuel oil required =16115 = 49.1%

From simulations, process energy demand = 144.2328×10^9 BTU/CD
= 22714 BPCD FOE

∴ Offsites = $\frac{32829 - 22714}{32829}$

= 30.8%

QUEBEC

Extra process energy required = 14921.16×10^6 BTU/CD

Base case process energy = 144.2328×10^9
incremental = 14921.16×10^6

∴ new demand = 1.5915×10^{11} = 25064 BPCD FOE

Offsites energy remains unchanged @ 10115 BPCD FOE

Hence new total refinery fuel req'd = 35179 BPCD FOE

Fuel gas production from simulations = 27756 BPCD FOE
LPG Sales 3586
24170 BPCD FOE

∴ FG = 68.7%
OIL = 31.3%

SECTION 7

7.0 LITERATURE REFERENCES

The following reference literature was used during the study:-

Ref. 1 - Petroleum Refineries In Canada, December 1979 (by Energy, Mines and Resources Canada)

Ref. 2 - Energy And Automotive Emissions, January 1980 (pre-print by International Lead Zinc Research Organization, Inc., New York)

Ref. 3 - Petroleum Refinery Engineering, W.L. Nelson Fourth Edition

Ref. 4 - Guide To Refinery Operating Costs, W.L. Nelson Third Edition

Ref. 5 - Petroleum Refining. Technology and Economics Gary and Handwerk

Ref. 6 - Lead Additive Consumption In Canada In 1979 - information supplied by Environment Canada

Ref. 7 - Canadian Refinery Receipts System, National Energy Board, 1980-04-17, printout of computer programme CRR060

Ref. 8 - Refined Petroleum Products, Statistics Canada Catalogue Number 45-004, December 1979

Ref. 9 - Oil And Gas Journal, April 15, 1963

Ref. 10 - A Guide To World Export Crudes, published by Oil and Gas Journal, 1976

Ref. 11 - - information supplied by Environment Canada (Sulphur levels)

Ref. 12 - Oilweek, June 9, 1980

Ref. 13 - Technical Data Book - Petroleum, Refining, American Petroleum Institute, Second Edition, 1970

Ref. 14 - ASTM STANDARDS - Classification of Diesel Fuel Oils (D975-537), Appendix II: Calculated Cetane Index

Ref. 15 - Hoffman, H.L., Hydrocarbon Processing, February, 1980

Ref. 16 - Eastman, Industrial and Engineering Chemistry, 1941

Ref. 17 - Lukenbach, E.C., Chemical Engineering Progress, February, 1979

Ref. 18 - Magee, J.S., R.E. Ritter and L. Rheaume, Hydrocarbon Processing, September, 1979

Ref. 19 - Adams, F.H. and A.P. Krueding, "Gasoline Production", Presented at the UOP Technology Conference, 1973

Ref. 20 - "Polygas Spells Relief From Alkylation Ills." Chemical Engineering, June 16, 1980

APPENDICES

ANNEX A

REFINERY CONSTRUCTION COSTS

Purpose

The purpose of this consulting contract is to determine the incremental costs to those companies in Canada which refine crude oil into gasoline, resulting from reducing the lead additive (tetra-ethyl lead and tetra-methyl lead) in gasoline from its current maximum level of 3.5 grams per imperial gallon (.77 grams per litre) to a lower level.

Background

Lead in the environment has been a longstanding concern because of its deleterious effects on human health. Children are more susceptible than adults to the unhealthy effects of lead. About 65% of total atmospheric lead comes from the combustion of leaded gasoline. Approximately 93% of Canadians are exposed to three quarters of this atmospheric lead. Consequently, there is a need to control the amount of lead entering the atmosphere from this source.

In 1979, thirteen companies operated about thirty-seven petroleum refineries which produced in Canada approximately 235 million barrels of gasoline, of which 30% was lead-free and 70% was leaded. Their refining capacity is broken down as follows:

Type of Unit	Feed Stock Capacity Barrels per day
Crude Oil Distillation	2,216,700*
Catalytic Cracking	598,900
Catalytic Reforming	413,700
Hydrogen Treating	983,600
Alkylation Units	58,300
Polymerization Units	21,350

*Also quoted as Crude Oil Refining Capacity

Canadian motor gasoline typically consists of the following components:

Component (Unit)	Volume, Million Barrels per year
Reformate (Catalytic Reforming Unit)	96
Catalytic Cracked (Catalytic Cracking Unit)	77
Alkylate*/Polymer* (Alkylation Units, Polymerization Units)	22
Light Naphtha (Crude Distillation Unit)	25
N-Butane	15
Total	235

*Feed stock from Catalytic Cracking Units.

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Tetra-ethyl lead (TEL) and tetra-methyl lead (TML) are added to motor gasoline to increase the octane number (research octane number) to 91 or better. This octane level would have to be achieved in other ways if TEL and TML could not be used as at present. In lead-free gasoline the octane rating is achieved through increased severity of the reformer operations and by increased application of alkylation and isomerization of components available from refinery streams but not suitable in their raw form for gasoline. Refineries not currently able to achieve 91 RON without TRL and TML would have to construct additional processing facilities.

An additional factor during the coming fifteen years will be that the change in the type of crude oil available could affect the refining process. For example, heavier crude and residual oil may comprise a much higher portion of refining feedstock and require additional facilities.

Reserves of oil are expected to decline in Canada. Consumption of gasoline is expected to be reduced during the forthcoming fifteen years because of both efficiency and conservation measures, resulting in the closing of some refineries. It is assumed that in the interim, these refineries may not wish to add processing facilities to accomodate the elimination of lead additives.

Terms of Reference

1. The consultant will consider three levels of control:

    a) Reduction in lead additives to gasoline to 2 grams per imperial gallon (.44 grams per litre).
    b) Reduction in lead additives to gasoline to 1 gram per imperial gallon (.22 grams per litre).
    c) Complete ban on lead additives to gasoline.

2. For each of the above levels of control, the consultant will determine the additional capital equipment which each of Canada's 36 refineries requires to accomodate the new levels of lead additive.

3. Using the determination in paragraph 2, the consultant will provide an estimate, for each refinery, of the capital costs which would be incurred to install the new processing equipment. The estimate will be in 1979 Canadian dollars.

4. The consultant will provide an estimate of the construction time required to install the new processing equipment. In making this estimate, the consultant will consider, in addition to other factors, the availability of construction employees, the backlog of equipment suppliers, the administrative time and engineering time required to launch these installations.

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5. The consultant will provide an estimate for each refinery of the additional number of employees necessary to operate the new processing equipment.

6. a) The consultant will provide, for each refinery, an estimate of the increase in feedstock necessary to yield gasoline at the required RON for the three levels of control.

   b) The consultant will also estimate for each refinery, any additional saleable by-products and any additional waste as a result of the change in product mix necessary to produce gasoline at the required RON. Using the above information, the consultant will estimate the "energy penalty" for each refinery.

   The energy penalty for producing a fixed quantity of lead-free gasoline is defined as the additional crude oil processed plus additional purchased power minus the energy equivalent of any additional liquified petroleum gas, or other product all divided by the quantity of crude oil required to produce an equal fixed quantity of leaded gasoline.

7. The consultant will provide an estimate of the increase in operating costs at each refinery for each level of control.

8. The consultant will provide an estimate of any increases in inventory and any additional working capital requirements which each refinery would need to operate the new processes.

9. Using the aforementioned information, the consultant will estimate the increase in 1979 Canadian dollars per litre of the costs of gasoline resulting from changes to the maximum allowable lead additive to leaded gasoline as specified in paragraph 1 of these terms of reference.

10. In order to adapt to reduced levels of lead in gasoline, petroleum refineries may be forced to abandon some of their equipment. The consultant will provide a general description of this equipment which each of the 36 refineries finds obsolete, and estimate its market value, (in 1979 dollars). When no market exists for the equipment, the consultant will estimate its scrap or salvage value.

11. As part of this study, the consultant will provide estimated conversion factors for each year up to 1985 which will convert the estimated construction costs in 1979 Canadian dollars to construction costs in subsequent years allowing for inflation or any other effect on the pricing of refinery construction.

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12. As part of this study, the consultant will provide opinions on the following topics. These opinions will be based on the consultant's knowledge of the industry. The consultant will not expend additional time researching the topics, or collecting information on them.

    a) The opportunities for specialization in the petroleum refining industry and in particular, the possibility of cost reductions if some refineries sell reformate to refineries unwilling to add capital equipment necessary to accomodate changes in lead additive. Such a strategy might be particularly suitable during a period of declining sales of gasoline.

    b) How the reduction in levels of lead additive to gasoline will affect the refining into gasoline of heavy crude (e.g. Lloydminster) and heavy residuals (e.g. bunker C).

    c) Probable annual gasoline production in Canada to the year 1995.

    d) Probable annual leaded gasoline production in Canada to the year 1995.

    e) The average hourly rates for additional employees required by the refineries to adapt to the three levels of control.

    f) The technological progress being made to refine gasoline to an adequate RON without using lead additives.

Notes:

1. All costs estimates will be in 1979 Canadian dollars.
2. Accuracy of estimated values is to be plus or minus 30% of actual values in 95% of the estimates.